PROCESS SIMULATION OF CRUDE OIL STABILIZATION SYSTEM : AN INDUSTRIAL CASE SUDY

by

MUHAMMAD FIRDAUS BIN DAINURE

Progress report submitted in partial fulfillment
of the requirements for the
Bachelor of Engineering (Hons)
(Chemical Engineering)

MAY 2013

Universiti Teknologi PETRONAS
Bandar Seri Iskandar
31750 Tronoh
Perak Darul Ridzuan

CERTIFICATION OF APPROVAL

PROCESS SIMULATION OF CRUDE STABILIZATION — AN INDUSTRIAL CASE STUDY

by

MUHAMMAD FIRDAUS BIN DAINURE

A project dissertation submitted to the

Chemical Engineering Programme

Universiti Teknologi PETRONAS

in partial fulfilment of the requirement for the

BACHELOR OF ENGINEERING (Hons)

(CHEMICAL ENGINEERING)

Approved by,

____

(Dr. Maziyar Sabet)

Date: 6 September 2013

UNIVERSITI TEKNOLOGI PETRONAS

TRONOH, PERAK

MAY 2013

CERTIFICATION OF ORIGINALITY

This is to certify that I, Muhammad Firdaus Bin Dainure (I/C No : 901109-04-5367), am responsible for the work submitted in this project, that the original work is my own except as specified in the references and acknowledgements, and that the original work contained herein have not been undertaken or done by unspecified sources or persons.

____

MUHAMMAD FIRDAUS BIN DAINURE

Student ID : 12731

I/C No : 901109-04-5367

Date : 06 September 2013

ABSTRACT

Petroleum is a naturally occurring flammable liquid of complex hydrocarbons found in oil reservoir beneath the Earth's surface. Crude oil (live crude) are sent to oil refineries for it to be processed into marketable petroleum products. However, before being transported to the refineries, the crude oil need to be stabilized beforehand since it contains light components that could flash off in low pressure conditions. This paper aims to find the suitable operating conditions to stabilize an incoming live crude feed to maximum True Vapor Pressure (TVPs) of 10 psia and 12 psia at Terengganu Crude Oil Terminal, TCOT. A simulations of the process has been conducted by using Aspen HYSYS (ver. 2007) software. It was found that at an heat exchanger outlet temperature of 85 - 90 ^{o}C, High Pressure Separator, V-220 A/B and Low Pressure Separator, V-230 A/B operating pressure of (400 - 592 kPa) and (165 - 186 kPa) respectively. The effects of major parameters, i.e. inlet feed properties, three phase separators operating pressure and pre-heaters trains performance on the product TVP are also studied. Based on the scenarios analyzed, it can be concluded that actual water volume (kbbl/d) has greater impact on heat exchanger's duty, thus incoming free water to TCOT should be less than 19.5 kbbl/d (9.1 vol%).

ACKNOWLEDGEMENT

Alhamdulillah, thanks to Allah S.W,T, whom with His willing giving me the opportunity to complete this Final Year Project.

First and foremost, I would like to express my deepest gratitude to my helpful supervisor Dr. Nejatollah Rahmanian, who has guided and support me during these two semester sessions to complete this project. The special thanks and gratitude also goes to my co-supervisor, Dr. Maziyar Sabet. The supervision and guidance that they gave truly help the progression and smoothness of my final year project. The co-operation is much indeed appreciated. I would also want to thank all lecturers and staffs of Chemical Engineering for their co-operations, suggestions and time responding to my inquiries along the way.

Not to forget, to all my friends and course mates for their cooperation, encouragement, constructive suggestion and full of support for this project completion, from the beginning till the end. Thanks to everyone who has been contributing by supporting my work during the final year project progress till it is fully completed.

Deepest thanks and appreciation to my beloved parents, and family for their love, support and prayers during my time completing this project. They also supported me and encouraged me to complete this task so that I will not procrastinate in doing it.

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

LIST OF ABBREVIATIONS

ABBREVIATIONS	DETAILS
TCOT	Terengganu Crude Oil Terminal
PCSB	PETRONAS Carigali Sdn. Bhd.
PGB	PETRONAS Gas Bhd.
GPP	Gas Processing Plant
LPGRU	Liquefied Petroleum Gas Recovery Unit
DPCU	Dew Point Control Unit
VRC	Vapor Recovery Compressor
TBC	Turbo Booster Compressor
CRS	Condensate Recovery System
GOR	Gas Oil Ratio
HX	Heat Exchanger
LP	Low Pressure
HP	High Pressure

CHAPTER 1

INTRODUCTION

1.1 BACKGROUND OF STUDY

Petroleum is a mix of naturally organic compound contains primarily hydrogen, carbon and oxygen. There are two types of petroleum which come straight out of the ground in form of liquid. The first one is called as crude oil and the second one is called as condensate. Crude oil is a dark and viscous liquid and condensate is a clear and volatile liquid. Crude oil usually in black color but it also comes in other colors like green, red or brown but they are not uncommon. Crude oil is come into two characteristics either it is light or heavy. Their characteristics depend on how this crude oil is vaporizing when it is heating, or it is added by chemical agents. It is light if it is a volatile oil and is heavy if it is viscous. A significant portion of the world crude oil is produced in the form of emulsion, R. C. Buruss & R. T. Ryder, (2003) [1].

The fluids existing in a hydrocarbon are usually under high pressure and may be in a liquid or gaseous state. The hydrocarbon fluids in the reservoir are usually in contact with water, which is normally salty. Each reservoir is unique : and all of its individual characteristics, listed below, will have an effect on how the fluid will be produced and how they must be treated when they reach the surface, M. George, (2010) [2].

a. Pressure
b. Temperature
c. Density of the fluids
d. Type and quantity of fluid that it contains
e. Whether the fluid contains components considered to be undesirable (e.g., hydrogen sulphide, H_2S and carbon dioxide, CO_2)
f. Amount of free water in the crude

The hydrocarbon fluids to be found in the reservoir range in a composition from nearly pure methane gas to a crude so heavy that it is essentially asphalt. The job of a production facility (oil & gas production terminal) is to separate the well stream into three components, typically called 'phases' (oil, gas and water), and process these phases into some marketable product(s) or dispose them in an environmentally acceptable manner. In a typical onshore oil and gas processing facility, separation is the heart of the process. Separation is a process to separate the gas from the liquid and the water from the oil. This is usually done in a separator which is an unfired pressure vessel. The well stream flows into the separator and allow the gas, oil and water to separate because the of the gravity. During the separations: crude oil stabilization process, gas is flashed from the liquids and 'free water' is separated from the oil. These step remove enough light hydrocarbons to produce a stable crude oil with the volatility (vapor pressure) to meet sales criteria. The gas that is separated must be compressed and treated for sales, while the free water produced will undergo further hydrocarbon recovery process as to reduce the oil proportion below 10 mg/L before being discharged as water effluents to the sea, Malaysia Environmental Quality Act, (1974) [9].

Separation is often accomplished in two or three stages of decreasing pressure : especially is this true in production from high-pressure wells. Staged separation is desirable for the following reasons as per below, Manning & Thompson, (1991) [5]:

a. It promotes more efficient separation. Less of the intermediate and heavier hydrocarbons is carried away by the gas : thus they remain in the liquid phase and a larger volume is retained. The liquid will also have a higher API gravity, increasing its value. Furthermore, less effort is required to remove the heavier hydrocarbons from the gas to condition it for pipeline transmission.
b. Less horsepower is required to compress the gas to pipeline pressure.

The existing separator design is capable of processing heavy crude oil and has been used widely over the world for so many years, however there is still need to improve the crude stabilization system as to achieve higher separation efficiency and cost effective.

1.2 PROBLEM STATEMENT

In oil & gas industry especially for oil & gas production facilities, managing oil-gas-water separation is an important and critical aspect as to treat the reservoir fluid to achieve standard storage/transport specification while maintaining a marketable quality. In order to maximize oil/gas production and increasing its market value, the oil and gas industry has shown keen interest in developing and optimize the separation efficiency between oil-gas-water in the crude stabilization process.

In addition, crude oil constituents must be reduced to an acceptable level determined by the purchaser. Table 1 shows the example of typical limit set by Exxon Pipeline Company:

Table 1: Exxon Mobil Exploration & Production, EMEPMI General Crude Specifications, (1985) [4]

Constituent	Limit	Remarks
Methane	100 ppm	by weight
Ethane	0.20%	by volume
Propane	1.00%	by volume
Hydrogen Sulphide	50 ppm	by weight
Carbon Dioxide	100 ppm	by weight
Basic Sediment and Water (BS&W)	1.00%	by volume
Inorganic Salts	75 lb per 1000 barrels	
Lead	0.010 ppm	by weight
Viscosity at Minimum Delivery Temperature	1000 SSU	216 centistokes
Reid Vapor Pressure	10 psia	
Maximum API Gravity	90^{o} API	
Minimum API Gravity	15^{o} API	

These are only intended as guidelines, because the true criterion is the local pipeline gauger acceptance (or Automatic Custody Transfer skid monitor), usually based on percent of Basic Sediment and Water (BS&W) in stabilized crude volume [4].

In reality, handling free water/produced water in crude is quite a challenging and critical aspect in crude stabilization process. High free water content (above 0.5% volume of BS&W set by PETRONAS Carigali Sdn Bhd) will eventually lead to the following operation losses:

a. Shipment demurrage due to poor crude quality.
b. Potential crude oil terminal plant slowdown/trip.
c. Potential gas compression system slowdown/trip.

Thus, it is crucial that the separation system works properly as to reduce the contaminant and undesirable constituents in the crude as well as to remove the produced water/free water in the crude oil receiving terminal.

In addition, due to the declining production of well fluid over the years, the associated petroleum gas (often called off-gas) also decreasing dramatically causing insufficient off-gas produced from the crude stabilization system in crude oil receiving terminal. This eventually lead to gas compressor surge and trip in the gas stabilization system. This can occur when the mass flow of gas to the compressor falls below a critical level with a high pressure difference across the machine. According to Norrie, (2010) [27] generally, if the suction flow drops too low, a 'Low-flow Trip' will shut down the machine. This 'Surging' in the machine can be very damaging to the compressor and associated piping and equipment due to heavy vibrations set up in the system. Also, surging can cause the machine to 'Overspeed' before the control system can react. This can also cause damage and is prevented by an 'Overspeed Trip Mechanism' which will again shut down the machine.

Frequent gas compressor surge/failure may results in production downtime or even unplanned plant shutdown. Off-gas produced from the stabilization train must be sent as sales gas to Gas Processing Plant (GPP) downstream for further treatment to be converted into valuable Liquefied Petroleum Product (LPG) or fuel gas for burners/heaters. Compressor failure will affect the sales gas production/transportation thus the remaining insufficient off-gas from crude stabilization train will be simply burnt off in gas flares or vented to atmosphere. The flaring of associated gas is controversial as it is a pollutant (CO_2), a source of global warming and is a waste of a valuable fuel source due to high content of hydrocarbons.

1.3 OBJECTIVES

The main aim of this project is to simulate a Crude Stabilization Unit with an inlet crude composition of Tapis Blend (Terengganu Crude Oil Terminal act as the receiving facilities) to obtain a stabilized crude with maximum True Vapor Pressure (TVP) of 83 kPa (12 psia) for storage/export. Thus, to accomplish the main aim, the following objectives need to be achieved:

1. To carry out process simulation (HYSYS) of Crude Stabilization System using Terengganu Crude Oil Terminal (TCOT) data as the case study.
2. To study on the effects operating conditions (Temperature, pressure, flowrate etc.) on the crude oil stabilization system.
3. To propose on the optimum inlet crude standard specification that will maximize the oil and gas production at crude oil processing facility.

1.4 SCOPE OF STUDY

This project will focus more on the crude stabilization system and identify any areas for improvement, thus the work/study will cover on the following activities as per below:

1. Study on the detailed process of Terengganu Crude Oil Terminal (TCOT) and carry out process simulation (HYSYS) as to observe/analyze the effect of changes in operating condition on oil & gas production.
2. Compare the existing TCOT crude stabilization system with typical oil production system.
3. Identify main critical equipment/facilities that affect separation efficiency.
4. Determine areas for improvement and design consideration to enhance separation efficiency.

CHAPTER 2
LITERATURE REVIEW

2.1 WELL FLUID COMPOSITIONS

Fluid flow from a well can include gas, free water, condensable vapors (water or hydrocarbon), crude oil, and solid debris (basic sediment). The proportion of each component varies in different well streams. When water is produced with crude oil, it is mixed in either or both of the following forms:

1. **Free Water/Produced Water**: Water mixed with the oil but will separate easily into a clear layer when the mixture is allowed enough time to settle.
2. **Emulsion**: Water can also be mixed with the oil in the form of very small droplets of water coated with oil. A mixture like this is called emulsion. Water in this case cannot be easily separated from oil and need to undergo demulsifications process in order to remove the water content in crude.

The waste water may be used as utility or discharged as water effluents to the sea. In either case, the water must be treated for solid particles removal, de-oxygenation, bacteria and hydrocarbon recovery at skimmer pit. As for the gas, it can be found in the well as [2]:

1. **Solution Gas**: Gas dissolved in the well fluids under the effect of pressure of the reservoir. As the fluids flow from the reservoir into the well and up to the surface, the pressure of the fluid decreases. The capacity of the liquid to hold gas in solution also decreases and gas starts to separate out of the oil.
2. **Free Gas**: Gas that is not held in the oil under reservoir conditions.
3. **Associated Gas**: Total gas produced with the oil in a crude oil well. This type of gas is separated in the three phase gravity separator and will under further treatment to be converted into Liquefied Petroleum Gas (LPG product), sales gas to petrochemical industries or used as a fuel gas for burner/heater at the production facility.

Non-hydrocarbon compound in crude may comprises of varying proportion of impurities depending on the characteristic of reservoir. Table 2 shows solid and gaseous impurities that may be produced with the crude oil:

Table 2: Impurities Content In Crude Oil, R. C. Buruss, (2003) [1]

Gaseous	**Solids**
Carbon Dioxide, CO_2	Asphaltene
Hydrogen Sulphide, H_2S	Wax
Carbon Disulphide	Sand
Nitrogen, N_2	Sulphur
Helium	Nickel, Vanadium
	Iron, Mercury

The quantities of the impurities are small but their presence may lead to reduction of efficiencies to the processing facilities such as the separator and compressor. Moreover, the presence of water and sediment leads to major difficulties such as corrosion, uneven heating, and plugging in heaters and exchangers and adverse effects on product quality. The level of water and insoluble impurities are usually measured as Base Sediment and Water (BS & W).

[4] Although crude oil assays evaluate various chemical properties of the oil, the two most important properties determining a crude's value are its density (measured as API specific gravity) and its sulphur content (measured per mass). Crude oil is considered 'heavy' if is high in wax content, or 'light' if low in wax content: an API gravity of 34 or higher is 'light', between 31-33 is 'medium', and 30 or below is considered 'heavy'. Crude may contain sulphur of varying quantities and are classified as follows:

a. Termed as **sour crude** when the sulphur content is more than (>2.5%/weight)
b. Termed as **sweet crude** when the sulphur content is less than (<0.5%/weight)

Crude oil produced in Malaysia has low sulphur content and high API gravity. Generally, the higher the API gravity (the 'lighter' it is), the more valuable the crude. It is of high quality and well sought after in the market. Table 3 shows the specifications of crude oil of certain countries:

Table 3: List and Specification of Crude Oil in Malaysia and other countries, Crude Oil Products, (2012) [7]

Product Name	API Gravity	Sulphur Content (as % of mass)	Field Location	Country
Bintulu Crude	36.6°	0.03%	Bintulu	Malaysia
Labuan Crude	32.0°	0.09%	Labuan	
Miri Crude	32.3°	0.08%	Miri	
Tapis Blend	45.2°	0.03%	Tapis	
Arab Heavy	27.7°	2.87%		Saudi Arabia
Arab Light	32.8°	1.97%		
Turkmen Blend	33.0°	0.15-0.29%	Aladzha, Okarem	Turkmenistan
Syrian Light	37.7°	0.74%	Banias, Tartous	Syria
Qatar Marine	35.8°	1.47%	Halul Island	Halul

2.2 GENERAL OIL HANDLING FACILITIES

2.2.1 Crude Processing & Treatment

Oil well fluids are produced normally in two phases - vapor and liquid. These two phases require entirely different handling, measuring and processing methods. Reservoir pressure are generally much higher than atmospheric pressure. As well fluids reach the surface, pressure on them is decreased. The liquid ability to hold gas in solution decreases and the liquids begin to release 'Solution Gas'. Light fluids begin to separate naturally when the pressure on them is lowered.

The solution gas released as 'Free gas' is held by the surface tension of the oil. Referring to Manning & Thompson, (1991) [5,] this free gas is released from the oil when the well fluids are warmed to reduce the surface tension of the oil. After the well fluid has been extracted from various oil well/well platform/subsea manifold, it will undergo primary separation to remove free water and any solid impurities (sand, wax etc.) at the surface production facilities (offshore platform).

Well fluid extracted from the well is received in the production manifold. Demulsifier chemicals is dozed in production manifold to promote breaking up of Water-Oil emulsion. Figure 1 shows the diagram of the flowline and manifolds :

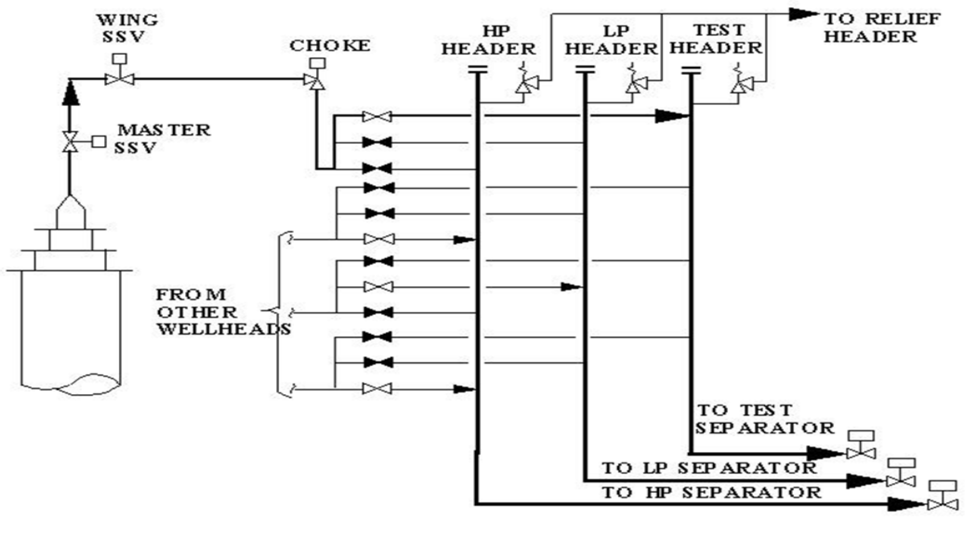

Figure 1: Diagram of Production Manifold [23]

Figure 2 (Attached in Appendix 1) shows the typical process for oil & gas processing plant, starting from the crude oil well and to the onshore crude oil terminal via offshore surface production platform.

K. Arnold & M. Stewart, (1991) [17] states that three phase well fluid is received from wells/well platforms and processed at large process platform generally consisting of the following four major processing module :

a. Separation (Oil, gas and produced water) & oil dispatch
b. Gas compression & dehydration
c. Produced water conditioning
d. Sea water processing & injection system

Many permanent offshore platform have full oil production facilities on board. Smaller platforms and subsea well must export raw production fluid to the nearest production facilities, which may be on a nearby offshore processing platform or an onshore terminal via pipeline for further treatment prior to storage or export Hazard Devold, (2006) [20]. Special corrosion inhibitor might be injected into the pipeline stream as to avoid the waxy formation and corrosion inside the pipeline.

Upon reaching onshore, the inlet crude will undergo further separation process (crude stabilization process) as to separate the oil mixture into three phase (oil-gas-water) streams. Most terminals have a series of production separators, starting with a high pressure separator, which separates the (HP) gas from the liquids. Liquids are then piped to a medium pressure (MP) separator, which removes more gas and the passes the liquids to a low pressure (LP) separator that removes even more gas and then separates water from the oil.

The produced water from the crude stabilization system is piped to a skim tank or to a drain pit, with the oil being piped to a metering and pumping station to be piped to other processes or storage tanks. Crude oil separation depend on the composition of the fluids, and on their pressure and temperature. The pressure of the fluids is controlled by the back - pressure regulator and the temperature may be regulated by expanding the fluid through a choke, by addition of heat in a furnace or by heating or cooling in a heat exchanger. Therefore, separators can be designed to handle fluids according to the fluid composition. The main principle used to achieve physical separation of gas and liquids are : **Gravity Settling** and **Coalescing.** Any separator may employ one or more of these principle, but the fluid phases must be 'immiscible' (cannot mix), and have different densities for separation to occur.

In summary, these are variables which aid in the separation of a fluid stream, SKG16 Facilities Process, (2010) [23]:

a. Temperature of the fluids
b. Pressure on the fluids.
c. Density of the components.

In addition to using the force of gravity, modern separators make use of other forces to get the best possible separation of oil and gas. The gas that is separated must be compressed and treated for sales. Compression is typically done by engine-driven reciprocating compressors while for large facilities or in booster service, turbine driven centrifugal compressors are used. Usually, the separated gas is saturated with water vapor and must be dehydrated to an acceptable level (normally less than 7 lb/MMscf) : this process is typically done in a glycol dehydrator. Dry glycol is pumped to the large vertical contact tower where it strips the gas off its water vapor. The wet glycol the flows through a separator to the large horizontal reboiler where it is heated and the water boiled off as a steam.

In overall, system capacity may be increased by [23]:

a. Fine tuning operating conditions of individual equipment items
b. Optimize process parameters or well fluid physical properties
c. De-bottlenecking production system

In some locations it may necessary to remove the heavier hydrocarbon to lower the hydrocarbon dew point : which is the minimum temperature where liquid might formed in gas phase. Contaminants such as H_2S and CO_2 may be present at levels higher than those acceptable to the gas purchaser and it is necessary to ‘sweeten’ the gas.

The oil and emulsion from the separators must be treated to remove water. Most oil contract specify a maximum percent of basic sediment and water (BS & W) that can be in the crude. This will typically vary from 0.5% to 3% depending on location. Some refineries have a limit on salt content in the crude, which may require several stages of dilution with fresh water and subsequent treating to remove the water. Typical salt limits are 10 to 25 pounds of salt per thousand barrels, Malaysia Environmental Quality Act, (1974) [9].

2.3 CRUDE STABILIZATION SYSTEM (MULTISTAGE SEPARATION)

Dissolved gas in the well fluid must be removed to meet pipeline, storage or tanker Reid Vapor Pressure (RVP) specifications. When the oil is essentially free of dissolved natural gas, then it can be stored in a vented tank at atmospheric pressure. The presence of the most volatile hydrocarbons (C1, C2, C3 etc.) increase RVP drastically. Removal of these dissolved natural gas components is called 'crude oil stabilization'. Crude oil can be stabilized by passing it through a series of flash drums or vapor-liquid-separator vessels at successively lower pressure. Stabilization can also sweeten the crude because the sour contaminant, H_2S, has a boiling point of -76.5 ^oF, intermediate to that of ethane and propane, Malaysia Environmental Quality Act, (1974) [9].

2.3.1 Effect of Separator Operating Pressure On Liquid Recovery

Because of the multi-component nature of the produced fluid, the higher the pressure if at which the initial separation occurs, the more liquid will be obtained in the separator. This liquid contains some light component that will vaporize in the storage tank downstream of the separator. If the pressure for initial separation is too high, too many light components will stay in the liquid phase at the separator and be lost to the gas phase at the tank condition (RVP : 12 psi). If the pressure is too low, not as many of these light components will be stabilized into liquid at the separator and they will be lost to the gas phase.

The tendency of any one component in the process stream to flash to the vapor phase depends on its partial pressure. The partial pressure of a component in a vessel is defined as the number of molecules of that component in vapor space divided by the total number of molecules of all components in the vapor space times the pressure in the vessel. Thus, if the pressure in the vessel is high, the partial pressure for the component will be relatively high and the molecules of that component will tend toward the liquid phase. Ken Arnold & Maurice Stewart (1999) [17] point out that as the separator pressure is increased, the liquid flow rate out of the separator increase.

Figure 3 shows the effects of separator pressure on the stabilized oil produced from the crude stabilization system:

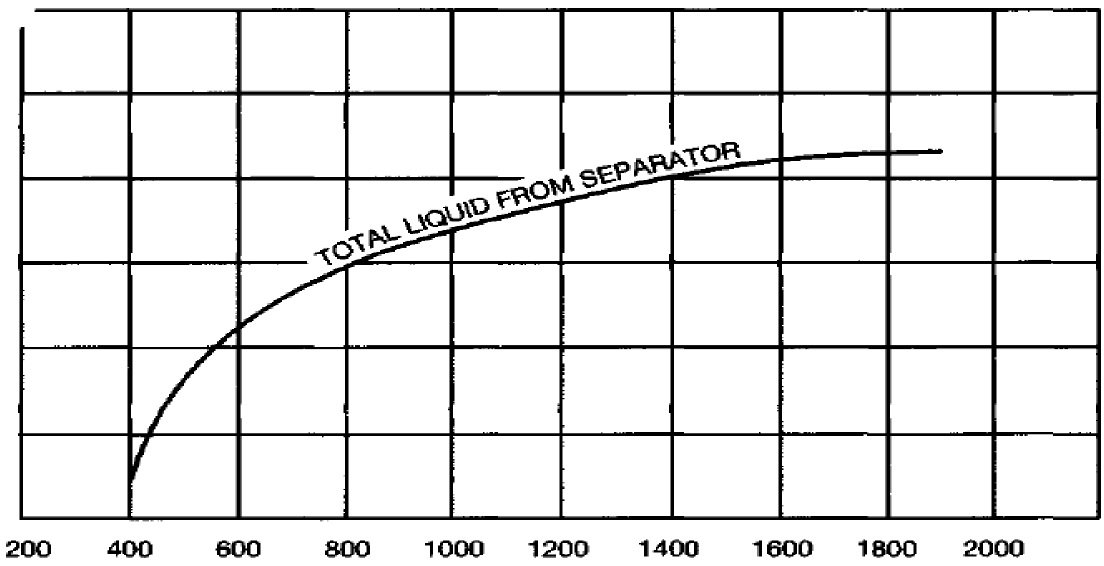

Figure 3: Effects of Separator Pressure on Stabilized Crude Production [17]

Based on figure 3, the total liquid produced from the separation increase with an increment of the separator's pressure. The problem with this is that many of these molecules are the lighter hydrocarbons (methane, ethane and propane), which have a strong tendency to flash to the gas state in atmospheric pressure conditions (storage). In the storage tank, the presence of these large number of molecules creates a low partial pressure for the intermediate range hydrocarbons (butanes, pentane, and heptane) whose flashing tendency at storage condition is very susceptible to small changes in partial pressure.

Thus by keeping the lighter molecules in the feed to storage tank we manage to capture a small amount of them as liquids, but we lose to the gas phase many more of the intermediate range molecules. That is why beyond some optimum point there is actually a decrease in storage tank liquid by increasing the separator operating pressure.

2.3.2 Effect of Number of Stages on Liquid Recovery

Because of the multi-component nature of the produced fluid, it can be shown that the more stages of separation after the initial separation the more light components will be stabilized into the liquid phase. In a stage separation process, the light hydrocarbon that flash are removed at relatively high pressure, keeping the partial pressure of the intermediate hydrocarbons lower at each stage. As the number of stages approach infinity, the lighter molecules are removed as soon as they are formed and the partial pressure of the intermediate components is maximized at each stage. The compressor horsepower required is also reduced by stage separation as some of the gas is captured at a higher pressure during separation process.

Ken Arnold & Maurice Stewart (1999) [17] states that as more stages are added to the process there is less and less incremental liquid recovery. The diminishing income for adding a stage must more than offset the cost of additional separator, piping, controls, space and compressor complexities. Usually, for each facility there is an optimum number of stages. In most cases, the optimum number of stages is very difficult to determine as it may be different from well to well and it may change as the well's flowing pressure declines with time.

Figure 4 shows an approximate guided to the number of stages in separation, which field experience indicate is somewhat near optimum:

Initial Separator Pressure, psig	Number of Stages
25 – 125	1
125 – 300	1 — 2
300 – 500	2
500 – 700	2 — 3

Figure 4: Stage Separation Guidelines [17]

2.3.3 Separator Operating Pressure

The choice of separator operating pressures in multi-stage system is large. The higher the operating pressure the smaller the compressor needed to compress the flash gas to sales. Compressor horsepower requirements are a function of the absolute discharge pressure divided by the absolute suction pressure. Increasing the low pressure separator may decrease the compression horsepower, however it may also add backpressure to inlet crude feed stream, restricting their flow, and allow more gas to be vented to atmosphere at the tank.

For practical reasons, the choice of separator operating pressures should match closely and be slightly greater than the compressor interstage pressures. The most efficient compressor sizing will be with a constant compressor ratio per stage. Therefore, an approximation of the intermediate separator operating pressures can be derived from:

$$\mathbf{R} = \left[\frac{\mathbf{Pd}}{\mathbf{Ps}}\right]^{1/n}$$

Where:

R= Ratio per stage

P_d= Discharge pressure, psia

P_s= Suction pressure, psia

n= Number of stage

Once a final compressor selection is made, these approximate pressures will be changed slightly to fit the actual compressor configuration.

2.4 TERENGGANU CRUDE OIL TERMINAL, TCOT OVERVIEW

Terengganu Crude Oil Terminal (TCOT) was commissioned in 1983, and has been operating for more than 25 years. It is located at East Peninsular Malaysia. TCOT's main function is to receive, stabilize, dehydrate, store and export crude oil from Terengganu's offshore fields. Dehydrated crude oil from TCOT is exported to customers through its export facilities such as SALM 1/SALM or to PETRONAS Penapisan (Terengganu) Sdn Bhd, TCOT Operating Manual, (1985) [22].

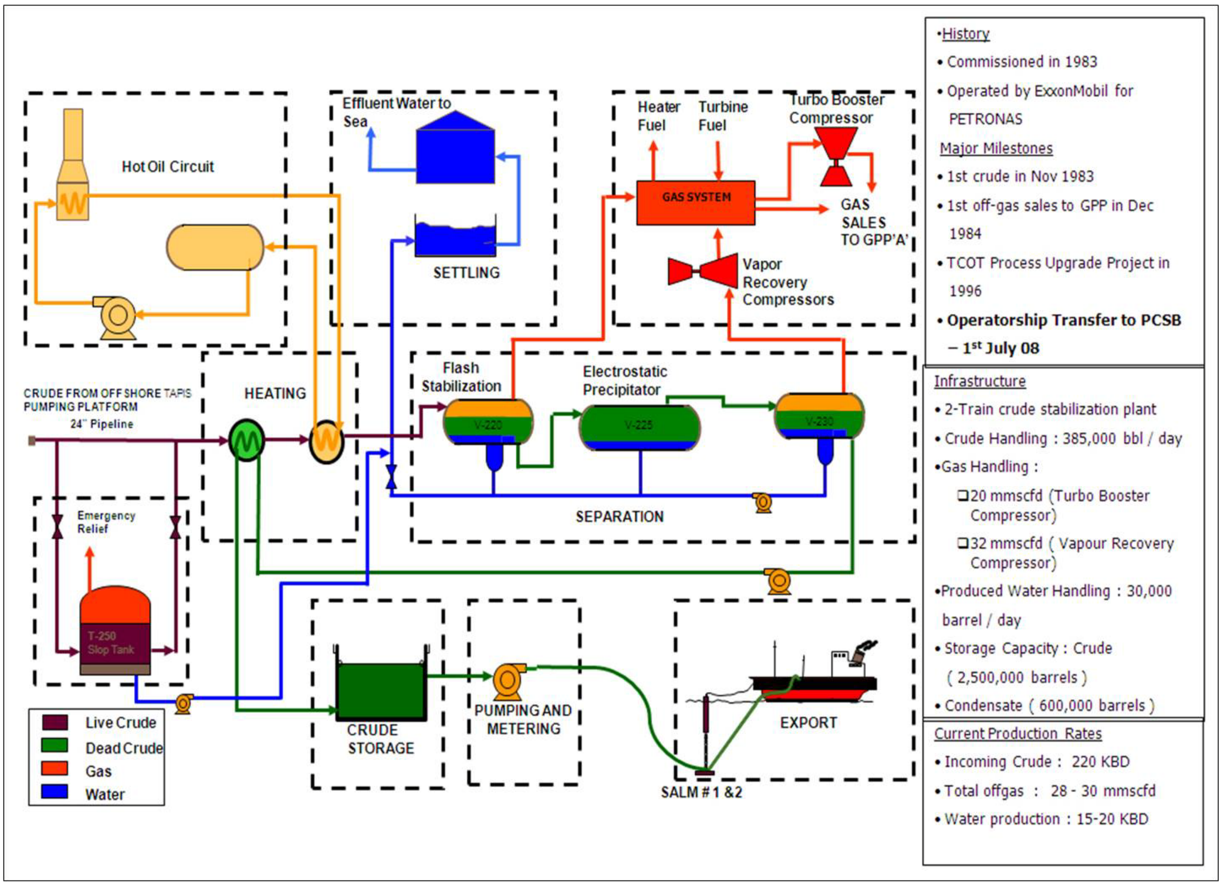

Figure 5: TCOT General Overview & Process

Off-gas from the crude stabilization is metered and exported to GPP1 Liquefied Petroleum Gas Unit (LPGRU) via Discharge Separator (V-272). Previously when the crude production was high, TCOT off-gas was sent to GPP1-LPGRU and the balance was sent to other GPP-DPCU in GPP-A through the Turbo Booster Compressor (TBC) and Vapor Recovery Compressor (VRC). Currently, the off-gas has decreased in volume as per decrease in oil production.

The TBC was initially operated when the LPGRU was unavailable. However, the TBC is currently obsolete, causing off-gas flaring in TCOT.

GPP-A has four Gas Processing Plants (GPP-1, 2, 3 & 4) which are currently receiving feed gas from Onshore Gas Terminal (OGT), Onshore Slug Catcher (OSC) and TCOT off-gas (mainly for GPP-1 only). The GPP-1 plant is divided into 2 plants, which are Dew Point Control Unit (DPCU) and Liquefied Petroleum Gas Recovery Unit (LPGRU). DPCU normally receives feed gas from OSC and OGT while LPGRU receives feed off-gas from TCOT. The LPGRU design capacity is 60 Tonne/hr (32 MMSCFD with gas molecular weight of 38) with minimum and maximum pressure of 300 kPa and 600 kPa respectively.

2.4.1 TCOT Crude Stabilization System

Based on TCOT Operation Train Manual, (2010) [24], TCOT crude stabilization system consist of two (2 x 50%) stabilization trains which each train is designed at 30, 605 m^3/d (1922, 500 bpd) with a total of two train maximum of 385, 000 bpd. The inlet pressure & temperature is designed at 1250 kpag & 27-29 ^oC. Each train has an operating valve (MOV-126 & MOV-127) for normal startup and shut down. Considering the factor of safety, each train has an inlet shutdown valve (SV-132 & SV-133) which activated on either abnormally high level or high pressure. TCOT crude stabilization system consist of several main equipment and are described as per below (Please note that detailed diagram for the equipment are attached in Appendix 2:

1. **Crude / Crude Exchangers (HX-210s)**
   a. 8 HX-210s crude to crude heat exchangers per train.
   b. Arranged in two parallel banks of four exchangers in series.
   c. Warm the inlet crude from 27-29 ^oC to 52-56 ^oC and at the same time cools the stabilized crude to about 40 ^oC before it is transferred to storage.

2. **Hot Oil / Crude Exchangers (HX-220s)**
   a. 2 parallel HX-220s hot oil to crude heat exchangers per train.
   b. One downstream of each parallel bank of four HX-210s.
   c. The HX-220s supply additional heat to 80 ^oC (normal operating)

3. **Gas-Oil-Water Separators (V-220s)**
    a. V-220 A/B operate as a three phase separator.
    b. Liquid retention time : 2 mins
    c. Gas flow : 27746 sm3/h each (23.5 mscfd)
    d. Liquid flow : 1488 m3/h each (224.7 kbpd)
    e. Pressure (operating / Design) : 443 / 724 kpag
    f. Temperature (Operating / Design) : 80 / 122 °C

4. **Electrostatic Separators (V-225s)**
    a. V-225 A/B operate as a two phase separator.
    b. Water content at outlet : 0.2 % volume
    c. Crude flow : 1407 m3/h each (212.5 kbpd)
    d. Water flow : 66 m3/h (10, 000 bpd)
    e. Pressure (operating / Design) : 313 / 724 kpag
    f. Temperature (Operating / Design) : 80 / 122 °C

5. **Crude Surge Separators (V-230s)**
    a. V-230 A/B operate as a three phase separator.
    b. Liquid retention time : 5 mins
    c. Gas flow : 15512 sm3/h each (13.1 mscfd)
    d. Liquid flow : 1353 m3/h each (204.3 kbpd)
    e. Pressure (operating / Design) : 130 / 552 kpag
    f. Temperature (Operating / Design) : 80 / 122 DegC

6. **Crude Oil Circulation Pumps (P-210s)**
    a. Three pumps per train
    b. Vertical canned-type pump
    c. Flow rate : 451 m3/h each (68.1 kbd)
    d. Differential heads : 137 m
    e. Rated KW : 160 KW
    f. SG : 0.743

2.5 BASIS OF STUDY (Tapis Blend – TCOT Inlet Crude Composition 2011)

This study will focus on Tapis Blend crude production, thus the inlet fluid composition data of Terengganu Crude Oil Terminal (TCOT) will be used as the basis of process simulation. Table 4 shows the summary of composition analysis of incoming feed to Terengganu Crude Oil Terminal, TCOT :

Table 4: TCOT Inlet Crude Feed Compositions

Components	Mol Fraction	Components	Mol Fraction
Hydrogen	0.0000	Undecanes	0.0471
Hydrogen Sulphide	0.0000	Dodecanes	0.0414
Carbon Dioxide	0.0051	Tridecanes	0.0423
Nitrogen	0.0003	Tetradecanes	0.0380
Methane	0.0414	Pentadecanes	0.0405
Ethane	0.0250	Hexadecanes	0.0292
Propane	0.0400	Heptadecanes	0.0244
i-Butane	0.0246	Octadecanes	0.0242
n-Butane	0.0292	Nonadecanes	0.0191
Neo-Pentane	0.0001	Eicosanes	0.0156
i-Pentane	0.0297	Heneicosanes	0.0137
n-Pentane	0.0211	Docosanes	0.0119
Hexane	0.0501	Tricosanes	0.0103
M-Cyclo Pentane	0.0134	Tetracosanes	0.0092
Benzene	0.0025	Pentacosanes	0.0081
Cyclo-hexane	0.0099	Hexacosanes	0.0072
Heptane	0.0463	Heptacosanes	0.0067
M-C-Hexane	0.0264	Octacosanes	0.0060
Toluene	0.0117	Nonacosanes	0.0058
Octanes	0.0591	Tricontanes	0.0051
E-Benzene	0.0033	Hentriacontanes	0.0044
M/P-Xylene	0.0181	Dotriacontanes	0.0034
O-Xylene	0.0048	Tritriacontanes	0.0029
Nonanes	0.0438	Tetratriacontanes	0.0022
1,2,4-TMB	0.0067	Pentatriacontanes	0.0019
Decanes	0.0494	Hexatriacontanes Plus	0.0179
		TOTAL	1.0000

Notes:

1. The fluid composition analysis are done based on dry basis mol fraction, TCOT Incoming Fluid Composition, (2011) [28].

2.6 DESIGN PRODUCTION RATE

Based on the composition and production profile, the facility design conditions are based on the conditions and parameter defined in Table 5 as per follows:

Table 5: Process Design Condition and Parameters

Parameters	Minimum Crude Production	Current Crude Production
Stabilized Crude		
Crude Production, kBd	60	195
Temperature, ^{o}C	40	40
True Vapour Pressure, psia	12	12
Water In Crude		
Temperature, ^{0}C	27	27
Pressure, Kpa (Abs)	1801	1801
Molecular Weight, g/mol	18.02	18.02
Dry Feed (Inlet Crude)		
Temperature, ^{0}C	27	27
Pressure, Kpa (Abs)	1801	1801
Molecular Weight, g/mol	180.57	180.57

Important notes:

1. Based on TCOT production profile obtained from TCOT Integrated Planning Department [25], TCOT can only sustain the minimum off-gas required within several years, while TCOT will reach its minimum design production rate. Thus, minimum crude production is expected at 60 kBD.
2. Stable operation at TCOT in which the flowrate of crude oil feed and the outlet temperature of Hot-Oil to Crude Exchanger (HX-220 AX-DX) are stable at designated temperature of (80 ^{o}C). Currently, TCOT are operating at 195 kBd crude production (As per year 2012 average crude production).

2.7 SELECTING THERMODYNAMIC MODEL

When faced with choosing a thermodynamic model, it is helpful to at least a logical procedure for deciding which model to try first. Elliot and Lira (1999) [23] suggested a decision tree as shown in Figure 6 as per below:

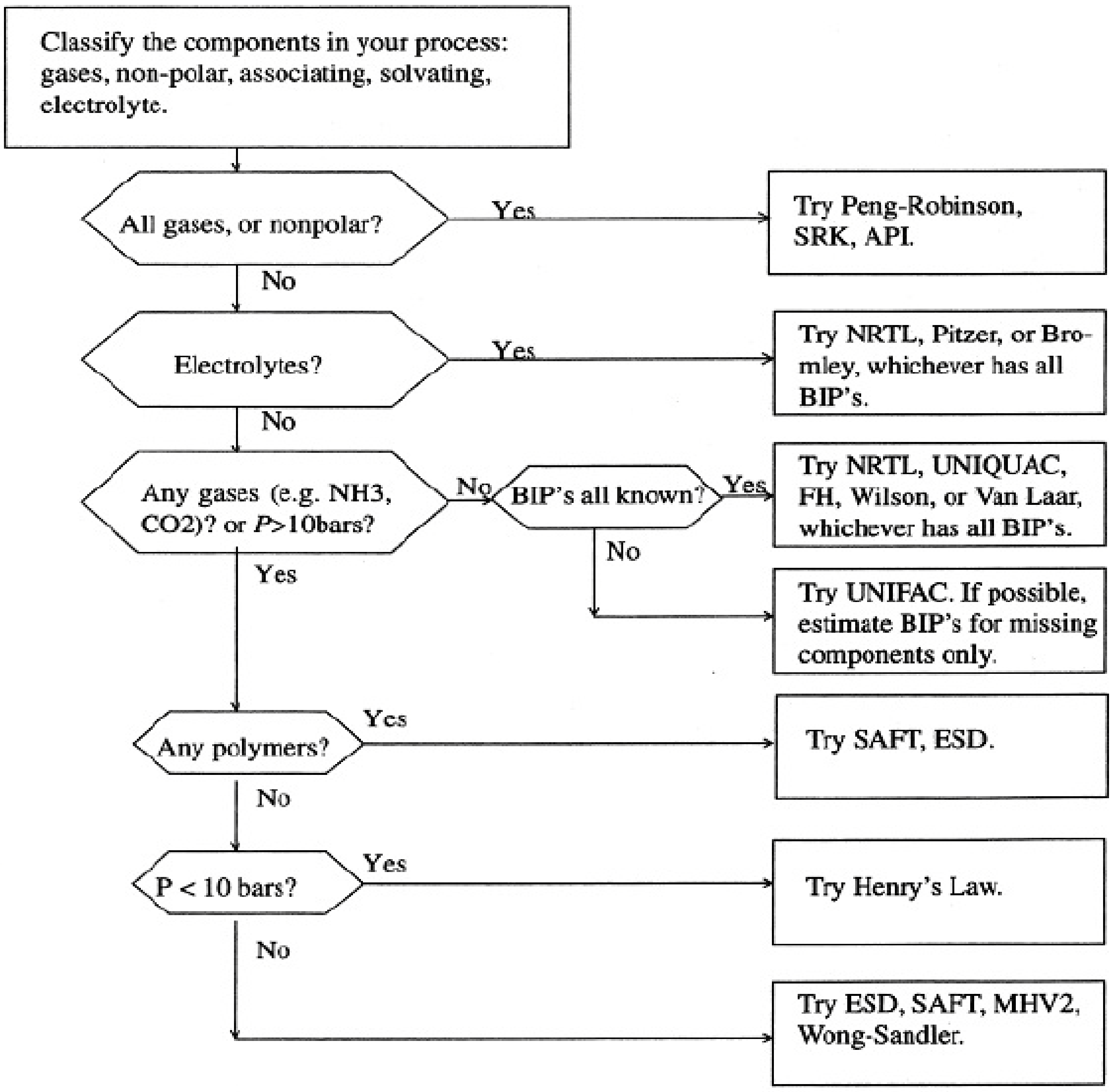

Figure 6: Thermodynamic Model Decision Tree

The property package available in HYSYS allow the user to predict properties of mixtures ranging from well defined light hydrocarbon systems as to complex oil mixtures and highly non-ideal (non-electrolyte) chemical systems.

HYSYS provides enhanced equations of state (Peng-Robinson, PR and PRSV) for rigorous treatment of hydrocarbon system: semi-empirical and vapor pressure models for the heavier hydrocarbon systems steam correlations for accurate steam property predictions: and activity coefficient models for the chemical systems. All of these equations have their own inherent limitations and the user are given with a wide choice of applications. Table 6 list some typical systems and recommended correlations:

Table 6: Typical System and Recommended Property Methods

Type of System	Recommended Property Method
TEG Dehydration	PR
Sour Water	PR, Sour PR
Cryogenic Gas Processing	PR, PRSV
Air Separation	PR, PRSV
Atm. Crude Towers	PR, PR Options, GS
Vacuum Towers	PR, PR Options, GS (<10 mmHg), Braun K10, Esso K
Ethylene Towers	Lee Kesler Plocker
High H_2 Systems	PR, ZJ or GS
Reservoir Systems	Steam Package, CS or GS
Hydrate Inhibition	PR
Chemical Systems	Activity Models, PRSV
HF Alkylation	PRSV, NRTL

For oil, gas and petrochemical applications, the Peng-Robinson EOS (PR) is generally the recommended property package to be used. Based on the Aspen HYSYS Property Wizard that helps the user to select the most appropriate property package for the simulations based on components and application, its recommended to use Peng-Robinson Fluid Packages as most of the components are hydrocarbons and non-polar (Equation of State, EOS). In addition, PR Fluid Packages is most enhanced in HYSYS, highest T & P range, has special treatment for key components, largest binary interaction database: good standards for hydrocarbons.

CHAPTER 3
PROJECT METHODOLOGY

3.1 GENERAL METHODOLOGY CHART

The chart below shows the general flow of this project from the beginning until the end:

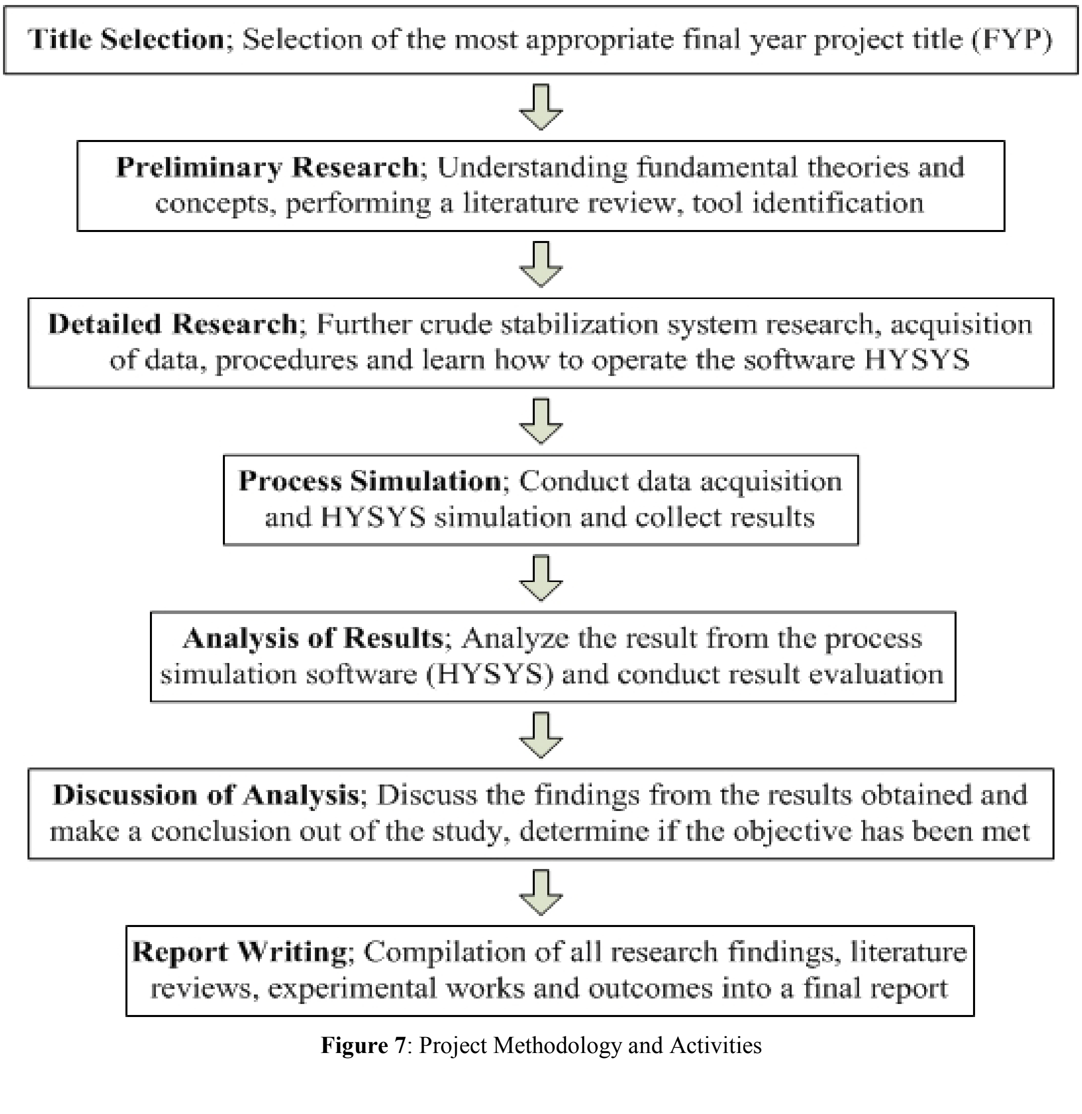

Figure 7: Project Methodology and Activities

For this project : Process Simulation of Crude Stabilization System, the methodology will be divided into two parts which are Project Research and Project Simulation.

3.2 PROJECT RESEARCH

In this part of the project, research on the topic of crude stabilization system is conducted by reading books, journals and article concerning the subject matter. Besides research on crude stabilization system itself, a brief background research is also done on the basis of crude oil terminal operation : in which for this study it is Terengganu Crude Oil Terminal (TCOT). From this research, it can be known why the crude stabilization is very important as to treat the well fluid according to the sales criteria. Besides, current technology being used is studied and compared between other crude oil receiving terminal.

3.3 PROJECT SIMULATIONS

Once thorough literature review and research has completed, detailed data acquisition on the process flow diagram of Terengganu Crude Oil Terminal (TCOT) as well as the estimates of operating conditions will be conducted. After all the required data has been collected, evaluation of data must be done as to analyze the reliability of each data. Then, the project simulation would be started by using ASPEN HYSYS® software. In this part, the process flow diagram of TCOT is generated using the software and the parameters inside the process will be adjusted as to obtain the desired/optimum results. The figure below show the general process simulation procedure that will be implemented in this research project:

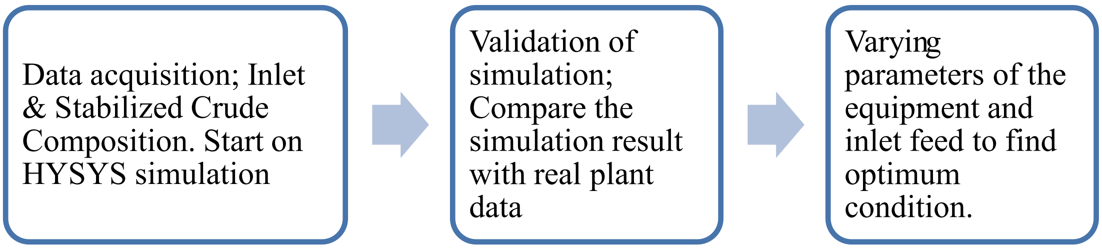

Figure 8: The schematic diagram depicting the general approach in this project

In this project, the process simulation will be more focused on Tapis Blend crude processing at TCOT. Thus, the inlet crude feed and stabilized crude composition were acquired from PETRONAS Carigali Sdn Bhd (PMO/PCSB). The simulation is intended as to find set of parameters/operating conditions that will optimize TCOT operation and crude oil production.

3.3.1 FYP1 Schedule and Gant Chart

Activities/Plan	Final Year Project 1 (FYP1)													
	1	2	3	4	5	6	7	8	9	10	11	12	13	14
Study/gather all data required for the simulation: Inlet & stabilized crude composition	■	■	■											
Study on Terengganu Crude Oil Terminal (TCOT) plant configuration and process : Inlet crude properties		■	■	■										
Evaluation of data in software and simulation : generation of case study and validation of operating data		■	■	■	■	■	■	■						
Identify critical equipment for crude stabilization system					■	■	■	■	■	■	■			
Finalize the best design for crude stabilization system using HYSYS (Based on year 2011 operating data)			■	■	■	■	■	■	■	■	■	■	■	■
Validation of simulation result with real plant data (TCOT) and result evaluation.									■	■	■	■	■	■
Key Milestones	**Final Year Project 1 (FYP1)**													
	1	2	3	4	5	6	7	8	9	10	11	12	13	14
Construction of process flow diagram of crude oil stabilization system.				■	■	■								
Validation of data collected							■	■	■					
Validation of HYSYS simulation with TCOT plant data & Site Visit to Terengganu Crude Oil Terminal (TCOT)										■	■	■	■	■

3.3.2 FYP2 Proposed Schedule and Gant Chart

Activities/Plan	Final Year Project 2 (FYP2)													
	1	2	3	4	5	6	7	8	9	10	11	12	13	14
Preliminary result and discussion	■	■												
Thorough analysis of the data generated and compare with the real plant data.			■	■	■									
Data analysis and interpretation based on the literature review and relate with plant application/situation.				■	■	■								
Detailed statistical analysis on off-spec data and conduct root cause analysis.					■	■	■	■	■					
Compile all data together and suggest an improvement in term of product's quality control.						■	■	■	■	■				
Recommend on optimum operating windows for crude oil stabilization operation.						■	■	■	■	■				
Finalize all data and result													■	■
Key Milestones	**Final Year Project 2 (FYP2)**													
	1	2	3	4	5	6	7	8	9	10	11	12	13	14
Thorough data analysis and interpretation	■	■	■	■	■									
Completion of FYP Final Report & Technical Paper						■	■	■						
Submission of Project Dissertation (Hard Bound)													■	■

CHAPTER 4
RESULTS AND DISCUSSIONS

4.1 BRIEF PROCESS DESCRIPTIONS

Figure 9 (Attached in Appendix 1) shows the process flow diagram of the simulation of the main crude stabilization unit using Aspen HYSYS (ver. 2006) software. HYSYS model are constructed based on the Piping & Instrumentation Diagram (P&ID) of Terengganu Crude Oil Terminal, TCOT [28]. Detailed material & energy balance are shown in Appendix 5 — HYSYS Simulation Report (Pseudo & Crude Assays).

It can be seen in the PFD, the main equipment governing the crude stabilization process is the staged separation using three phase separator (V-220 A/B, V-225 A/B, V-230 A/B). The inlet crude from offshore platform (TAPIS Pump) at maximum temperature and pressure of 17 barg and 27 °C are first heated via HX-210 A-R and HX-220 A-D to achieve required temperature for the staged separation process at 80 °C. The crude oil is stabilized and dehydrated in a crude stabilization process prior to being stored in floating roof storage tanks.

The crude stabilization facilities consist of two trains and each trains comprised of Crude to Crude Heat Exchangers (HX210 A-R), Hot Oil to Crude Exchangers (HX220 A-D), Gas-Oil-Water Separator (V220 A/B). Stabilized Crude Oil form Crude Surge Separators is piped through HX210 A-R to the storage tank. Stabilized crude specifications and operating target are defined as below:

Table 7: Stabilized Crude Specifications (Storage & Export Line)

Product	Specifications	Operating Target
Export Crude	0.5 % BS&W	< 0.3 % BS&W
	TVP : 83 kPa	
	Temp. : 38 °C	

Any malfunction of the stabilization train equipment that prevents completion of the stabilization process will cause the stabilization train to be blocked in and the inlet flow to the train will be diverted to Slop Oil Tank (T250) with a design capacity of 62 918 BBL (operating pressure and temperature of 13.8 kpag and 38 ^{o}C). Hydrocarbon Vapor from V230 A/B is sent directly to Discharge Separator (V272), while the hydrocarbon vapor from V230 A/B is scrubbed by suction scrubber (V270 A-D) and compressed via Vapor Recovery Compressor, VRC (C270 A-D) before sending to V272. The vapor from VRC is cooled by After Cooler (HX271 A-D) prior to send to V272. The combined gas form V272 is metered and sent to LPGRU.

The condensate formed in V272 is piped to LP Condensate Recovery Separator (V274) is sent to VRC suction Scrubber (V270 A-D) inlet while the condensate form V272 is sent to Atmospheric Condensate Recovery Separator (V277). The flashed gas from V277 is sent to LP flare. The condensate collected in V277 is metered and sent to condensate storage.

The off-gas from the crude stabilization system which are mainly from the high pressure separator (V-220 A/B) and low pressure separator (V-230 A/B) are gathered and sent to gas stabilization header. After the off-gas has been dried and free from any water and condensates, it will be sent as sales gas to Liquefied Petroleum Gas Recovery Unit (LPGRU) and Dew Point Control Unit (DPCU), located in GPP1 via gas compression system (Turbo Booster Compressor, TBC and Vapor Recovery Compressor, VRC.

4.2 PROCESS SIMULATIONS CASE STUDY

Based on the inlet crude compositions and production profile, there are two cases considered for this study, which are:

1. Crude Stabilization HYSYS model based on pseudo components
2. Crude assay (HYSYS Oil Manager).

Basis inlet compositions used for both case studies are shown in section 2.5 (Tapis Blend - TCOT Inlet Crude Compositions 2011). Both cases are simulated with a current stabilized crude production of 195 kBD. Detailed description for each case study and simulation validation are explained in further section as per below:

4.2.1 HYSYS Crude Stabilization Model (Pseudo Component)

The feed used for the simulation in this project was based on TCOT inlet Crude Composition from Tapis Pump Offshore (Please refer section 2.5). Heavy components from C31* to C36+* are lumped together to form new pseudo components. The new pseudo components properties are defined as per below [28]:

Table 8: Pseudo Components Properties

Pseudo Component	Molecular Weight	Ideal Liquid Density, kg/m³
C31*	435.3	813.9
C33*	464.3	816.3
C34*	478.3	816.7
C35*	492.4	817.1
C36+*	599.7	930.0

These pseudo components will be used along with the pure components from Aspen HYSYS (ver. 2006). Detailed inlet composition and feed properties used for this simulation are tabulated as per Table 9 & 10 below:

Table 9: Inlet Crude Composition HYSYS Pseudo Components

Components	Mol Fraction	Components	Mol Fraction
Hydrogen	0.000000	Dodecanes	0.018942
Hydrogen Sulphide	0.000000	Tridecanes	0.019353
Carbon Dioxide	0.002333	Tetradecanes	0.017386
Nitrogen	0.000137	Pentadecanes	0.018530
Methane	0.018942	Hexadecanes	0.013360
Ethane	0.011438	Heptadecanes	0.011164
Propane	0.018301	Octadecanes	0.011072
i-Butane	0.011255	Nonadecanes	0.008739
n-Butane	0.013360	Eicosanes	0.007137
Neo-Pentane	0.000046	Heneicosanes	0.006268
i-Pentane	0.013589	Docosanes	0.005445
n-Pentane	0.009654	Tricosanes	0.004713
Hexane	0.022922	Tetracosanes	0.004209
M-Cyclo Pentane	0.006131	Pentacosanes	0.003706
Benzene	0.001144	Hexacosanes	0.003294
Cyclo-hexane	0.004530	Heptacosanes	0.003065
Heptane	0.021183	Octacosanes	0.002745
M-C-Hexane	0.012079	Nonacosanes	0.002654
Toluene	0.005353	Tricontanes	0.002333
Octanes	0.027040	Hentriacontanes	0.002013
E-Benzene	0.001510	n-Dotriacontanes	0.001556
M/P-Xylene	0.008281	Tritriacontanes	0.001327
O-Xylene	0.002196	Tetratriacontanes	0.001007
Nonanes	0.020040	Pentatriacontanes	0.000869
1,2,4-TMB	0.003065	Hexatriacontanes Plus	0.008190
Decanes	0.022602	Water	0.542245
Undecanes	0.021549		1.000000

Important notes:

1. Water stream are added as a different stream with inlet crude stream (dry basis mole fraction) ranging about 10% of the inlet crude flow.
2. Detailed crude component and specifications are attached in the Appendices.

4.2.2 HYSYS Crude Stabilization Model (Crude Assay Compositions)

Aspen HYSYS (ver. 2006) are used to simulate the base Crude Oil Stabilization process using crude assay technique. Main aim is to characterize an oil using chromatographic data of incoming crude (Please refer section 2.5). The petroleum characterization method in Aspen HYSYS converts laboratory analyze of crude oils, petroleum cuts and etc. into a series of discrete hypothetical components. The data and compositions used to characterize the crude assay are divided into several categories which are tabulated as per Table 10, 11, 12 below:

Table 10: Detailed Compositions of Light Ends Components

Light Ends Components Groups		
Formula	**Components**	**Mol Fraction**
H2	Hydrogen	0.0000
H2S	Hydrogen Sulphide	0.0000
CO2	Carbon Dioxide	0.0051
N2	Nitrogen	0.0003
C1	Methane	0.0414
C2	Ethane	0.0250
C3	Propane	0.0400
i-C4	i-Butane	0.0246
n-C4	n-Butane	0.0292
Neo-C5	Neo-Pentane	0.0001
i-C5	i-Pentane	0.0297
n-C5	n-Pentane	0.0211

Light ends are defined as pure components with low boiling points. Components in the boiling range of C1 to n-C5 are categorized under light ends components. These light end play an important part in characterizing the oil cut. Below are the bulk properties of the crude which indicate average crude inlet properties.

Table 11: Bulk Properties of the Inlet Crude

Whole Sample Properties			
Phase	Liquid	Gas	Whole Fluid
Average Mole Weight (g/mol)	180.57	44.4	158.11
Measured density at 15.6 ^{o}C (g/cm³)	0.82		
Real Relative Density		1.56	
Entrained Water Content (wt%)	0.05		

Table 12: Detailed Composition of Incoming Crude to Characterize the Crude

Paraffinic Component Groups		
Formula	**Components**	**Mol Fraction**
C6	Hexane	0.0501
C7	Heptane	0.0463
C8	Octanes	0.0591
C9	Nonanes	0.0438
C10	Decanes	0.0494
C11	Undecanes	0.0471
C12	Dodecanes	0.0414
C13	Tridecanes	0.0423
C14	Tetradecanes	0.0380
C15	Pentadecanes	0.0405
C16	Hexadecanes	0.0292
C17	Heptadecanes	0.0244
C18	Octadecanes	0.0242
C19	Nonadecanes	0.0191
C20	Eicosanes	0.0156
C21	Heneicosanes	0.0137
C22	Docosanes	0.0119
C23	Tricosanes	0.0103
C24	Tetracosanes	0.0092
C25	Pentacosanes	0.0081
C26	Hexacosanes	0.0072
C27	Heptacosanes	0.0067
C28	Octacosanes	0.0060
C29	Nonacosanes	0.0058
C30	Tricontanes	0.0051
C31	Hentriacontanes	0.0044
C32	Dotriacontanes	0.0034
C33	Tritriacontanes	0.0029
C34	Tetratriacontanes	0.0022
C35	Pentatriacontanes	0.0019
C36+	Hexatriacontanes Plus	0.0179

Aromatic Components Groups		
Formula	**Components**	**Mol Fraction**
	Benzene	0.0025
	Toluene	0.0117
	E-Benzene	0.0033
	M/P-Xylene	0.0181
	O-Xylene	0.0048
	1,2,4-TMB	0.0067
Naphtenic Components Groups		
Formula	**Components**	**Mol Fraction**
	M-Cyclo Pentane	0.0134
	Cyclo-hexane	0.0099
	M-C-Hexane	0.0264

Important notes:

1. These composition will be specified under categories in Aspen HYSYS.
2. The crude assay cuts obtained by HYSYS are characterize by its boiling point.
3. The mole fractions are done on dry basis.

Based on the specified inlet compositions and properties, Aspen HYSYS (ver. 2006) will generate a hypothetical components according to the oil cut specified by users. Detailed composition of the crude cuts obtained from HYSYS are tabulated as per Table 13 below:

Table 13: Detailed Compositions of HYSYS Simulation (Crude Assay)

Components	Mol Fraction	Components	Mol Fraction
Hydrogen	0.000000	NBP[0]173*	0.029842
Hydrogen Sulphide	0.000000	NBP[0]188*	0.014135
Carbon Dioxide	0.002450	NBP[0]203*	0.015755
Nitrogen	0.000145	NBP[0]217*	0.015594
Methane	0.020018	NBP[0]232*	0.015599
Ethane	0.012088	NBP[0]246*	0.016323
Propane	0.019341	NBP[0]260*	0.016082
i-Butane	0.011895	NBP[0]275*	0.013129
n-Butane	0.014119	NBP[0]289*	0.011920
Neo-Pentane	0.000048	NBP[0]304*	0.012228
i-Pentane	0.014361	NBP[0]318*	0.011605
n-Pentane	0.010202	NBP[0]332*	0.008665
NBP[0]43*	0.011555	NBP[0]347*	0.007694
NBP[0]57*	0.010617	NBP[0]362*	0.007327
NBP[0]70*	0.012442	NBP[0]376*	0.006767
NBP[0]85*	0.021158	NBP[0]390*	0.005854
NBP[0]100*	0.025856	NBP[0]405*	0.005138
NBP[0]117*	0.023148	NBP[0]420*	0.004750
NBP[0]129*	0.015839	NBP[0]438*	0.007011
NBP[0]146*	0.020609	H20	0.516477
NBP[0]158*	0.012217	TOTAL	1.000000

Important notes:

1. Water are added as different stream along with the dry inlet feed to the crude stabilization section with liquid volume flow of 10% of dry inlet crude flowrate.

Table 14: Feed Stream Properties

Properties	
Vapor/Phase Fraction	0.0000
Temperature, ° C	27
Pressure, kPa Abs	1801
Molar Flow, kgmole/h	13229.58
Mass Flow, kg/h	1104875.71
Std Liquid Volume Flow, barrel/day	195500
Molecular Weight	83.52
Average Mass Density, kg/m3	778.90

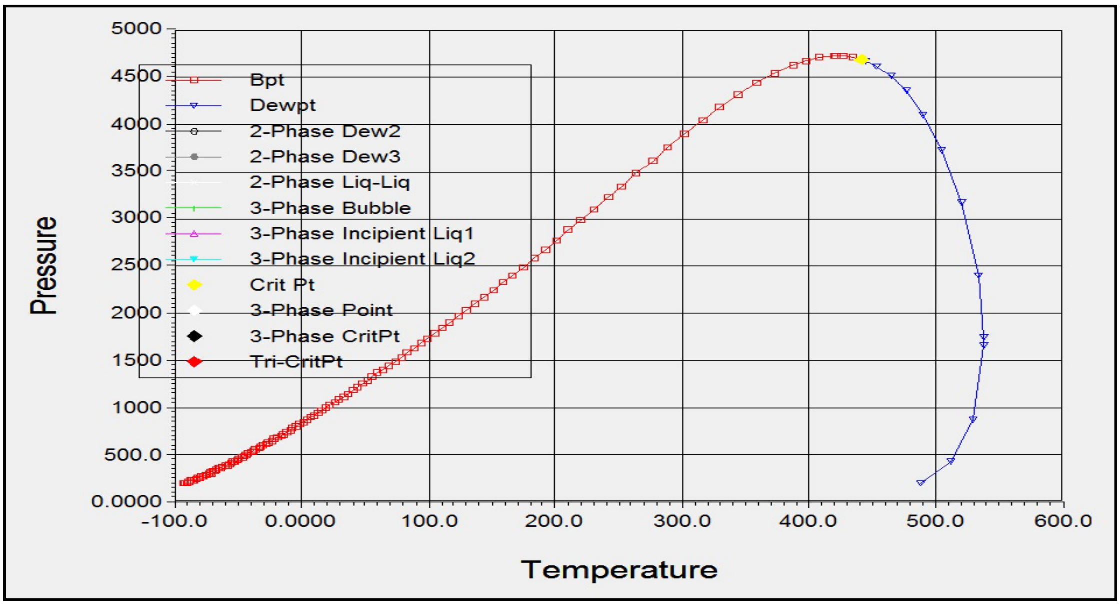

Figure 10: Phase Envelope Curve for Inlet Feed

The phase envelope diagram in Figure 10 shows the bubble points and dew points of the inlet crude at different pressures. The phase envelope was calculated by Aspen HYSYS on dry basis. According to Francis S. Manning & Richard E. Hompson (1995) [29], between the bubble point and dew point curves, the hydrocarbon is in two phase which is vapor–liquid. An obvious conclusion of this behavior is that a hydrocarbon mixture has a boiling range at constant pressure rather than a boiling point. The incoming feed from Tapis Pump (offshore platform) to Terengganu Crude Oil Terminal, TCOT is at 27 °C and 17 barg (1801 kPa Abs.) As can be seen based on the phase envelope above, the incoming feed will be in pure liquid state. The incoming live crude from offshore platform usually comprises of produced water and also contaminant or typically referred as Basis Sediments & Water (BS&W).

4.3 PROCESS SIMULATION VALIDATIONS

4.3.1 Terengganu Crude Oil Terminal Stabilized Crude Compositions

The stabilized crude compositions are obtained from Oil Movement Technologist Department, PETRONAS Carigali Sdn. Bhd (PMO/PCSB), Kerteh. Figure 11 shows a graph of stabilized crude compositions at Terengganu Crude Oil Terminal for year 2011. The sample was taken at the transportation line of the sales crude (after stabilization process).

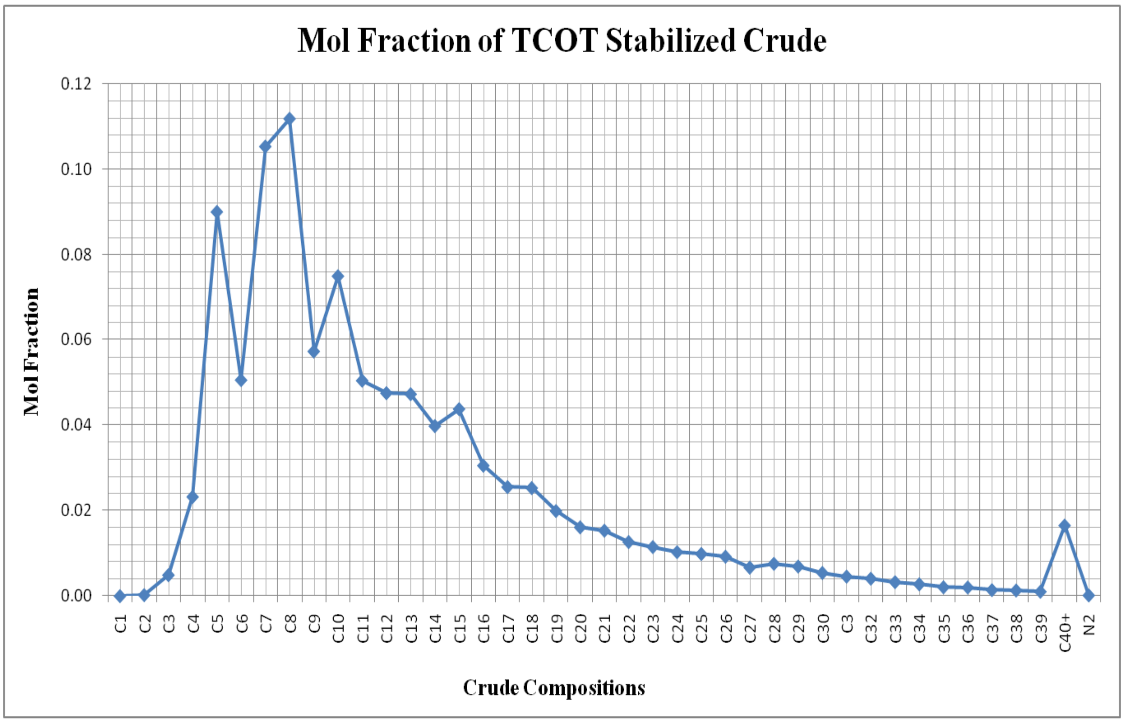

Figure 11 : Terengganu Crude Oil Terminal, TCOT Stabilized Crude Compositions

Based on figure 11, most of the volatile components such as methane and ethane and propane (C1 – C3) in the live crude have been flashed off during crude oil stabilization system under high pressure different in the oil, gas and water separator vessels. The off-gas produced in the crude stabilization area are collected and purified prior sending to the Gas Processing Plant, GPP as sales gas. The stabilized crude (dead crude) which have less amount of the volatile component now have lower vapor pressure which can be easily stored and transported. The True Vapor Pressure, TVP of the stabilized crude should not be more than 12 psia. Any higher than that, the outlet crude is off specification and will recycled back to the process trains.

In addition, crude oil constituents must be reduced to an acceptable level determined by the purchaser. Fortunately, Malaysisa's crude oil are mostly sweet crude which comprises low sulphur content and high API gravity. Generally, the higher the API Gravity (the lighter it is), the more valuable the crude. It is of high quality and well sought after in the market. Table 15 below shows the preliminary analysis on the stabilized crude properties.

Table 15 : Preliminary Analysis on Stabilized Crude at TCOT [30]

TEST	METHOD	UNIT	RESULT
Density@ 15 °C	ASTM D 5002	g/cm^3	0.8058
API Gravity @ 60°F	calculated	degree	44.1
Specific Gravity	calculated	-	0.8062
Total Sulphur	ASTM D 4294	wt%	0.0345
Nitrogen Content	ASTM D 664	ppm wt	216
Flash Point	IP 170	°C	<0
Pour Point	ASTM D 5853	°C	18
Reid Vapor Pressure @37.8°C	ASTM D 323	kPa	54.5
Salt Content	ASTM D 3230	lb/1000bbls	8.7
Kinematic Viscosity @40 °C	ASTM D 445	cSt	2.458
Gross Calorific Value	ASTM D 240	MJ/kg	45.755
Mercury	UOP 938	ppb wt	38
Basic Sediment & Water	ASTM D 4007	vol %	0.25
Water Content	ASTM D 4006	vol %	0

Based on Table 14, Tapis blend crude has high API gravity of 44.1 and considered as quality crude. On the other hand, the Reid Vapor Pressure of the stabilized crude is less than 12 psia (83 kPa) which meet the required specification for storage and transportation. The stabilized crude will be usually transported through pipeline or shipment. Furthermore, usually the live crude coming from the offshore platform is contaminated with sediment which includes sulphur (merchaptants, hydrogen sulphide etc.) and must be reduced within acceptable limit.

Crude oil stabilization system is designed to sweeten the live crude by flashing off the volatile component and reduce the amount of contaminants/sediments inside it. The live crude also will be dehydrate to remove any free water which might result in corrosion and give bad impact to the processing facilities downstream.

4.3.2 HYSYS Simulation (Pseudo Components) Validation

In order to ensure that the simulation done in this project is valid, the compositions of the final product obtained by HYSYS Simulations are compared to Terengganu Crude Oil Terminal (TCOT) stabilized crude compositions. Figure 12 below shows a graph of component's mole fraction (dry basis) vs. crude components. The two different trends represent two dif ferent stabilized crude composition which are the TCOT plant data and also stabilized crude compositions obtained from the Aspen HYSYS (ver. 2006) — Pseudo components analysis. The complete data used for the simulation validation in table form can be referred in Appendix IV.

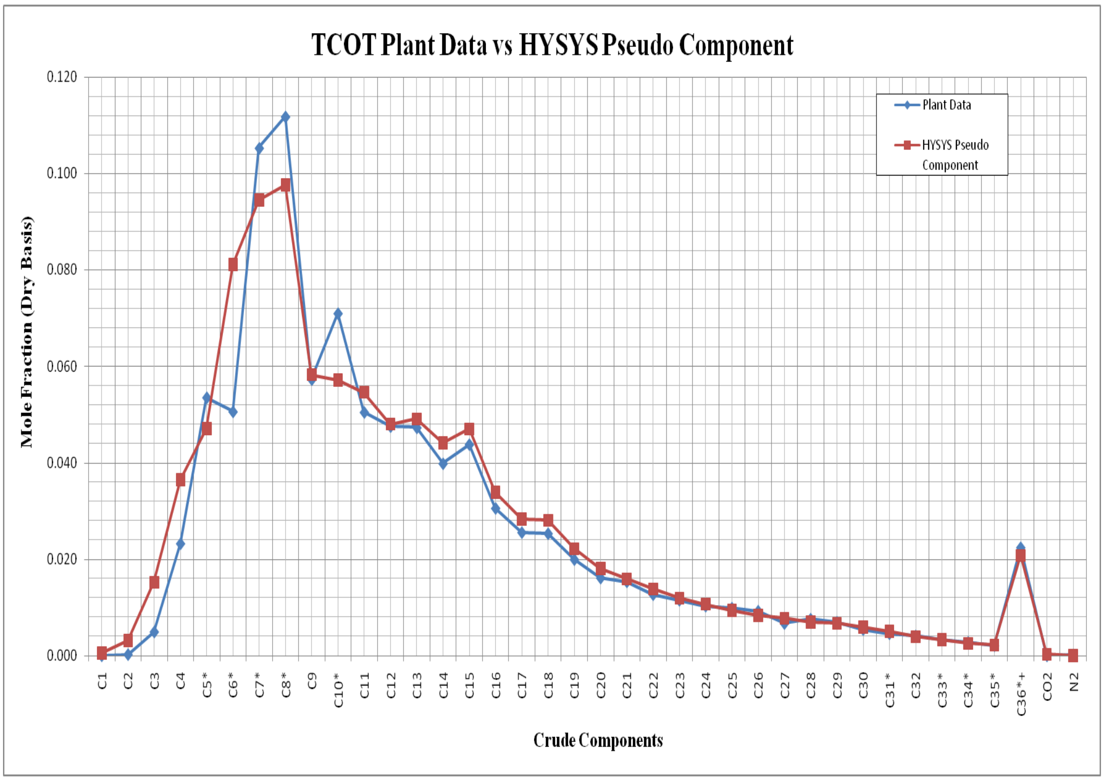

Figure 12: Comparison of Product Compositions

The mol fraction comparison are done on dry basis which ignore the composition of water in the crude. Detailed water content analysis for each simulation will be explained further in Section 4.4 — Result Analysis. Based on the data in Figure 11, it can be seen that the HYSYS Simulation results in high mol fraction of light component compared to the plant data. There are two components which shows high deviations in the comparison:

1. Hexane, C6
2. Decane, C10

Based on the crude compositional analysis, these components were unable to be quantified due to co-eluting with other components. Thus, the amounts were group together as lumped component. Total components of the lumped components are tabulated as per table 16 below:

Table 16: Detailed Compositions of Lumped Components of TCOT Plant Data

C5*	n-C5
	Cyclopentane
	2-Methylpentane
	3-Methylpentane
C6*	n-C6
	Methylcyclopentane
	2,4-dimethylpentane
	Benzene
	Cyclohexane
C7*	n-C7
	Methylcyclohexane
	Toluene
C8*	n-C8
	Ethylbenzene
	Meta & Para-Xylene
	Ortho-Xylene
C10*	n-C10
	n-Butylbenzene*

Based on the preliminary analysis, TCOT stabilized crude contains more intermediate components compared to the HYSYS. On the other hand, the composition of paraffinic component and heavy crude components from C11 to C36+* are almost the same for both plant data and the simulation. In addition, there are only trace amounts of Carbon Dioxide, CO_2 and Nitrogen, N_2. Due to absence of pure heavy component in HYSYS, pseudo components of C36+* were created (hypothetical components) which comprises of the following components:

Table 17: Detailed Components of Pseudo C36+*

C36+*	C36
	C37
	C38
	C39
	C40+

From the above analysis, it can be concluded that the HYSYS data contains the lightest components followed by the real plant data. Besides, TCOT plant data results in the crude the most intermediate components from C7* to C10*. In general, both stabilized crude compositions either from TCOT plant data and HYSYS simulation, produced crude with balanced heavy component from C11 to C36+*.

However, in overall, the trend of the mole fraction of the components is similar for all two sets of data. There are no major differences and thus, it is proven that the simulation done using the HYSYS software is valid and can be the basis of predicting tools for operational purpose.

4.3.3 HYSYS Simulation (Crude Assay) Validation

As per explained before, generally there are two methods to simulate the crude oil stabilization system using HYSYS simulation which is by:

1. **Pseudo Components**: Due to lack of pure component which is mostly represents heavy crude in HYSYS simulation; hypothetical components are created based on the crude characterization properties done by laboratories.
2. **Crude Assay**: Based on the crude sampling, the light component (C1 — C6) can be used to characterize the heavy component in the simulations. The heavy component are characterize based on the boiling points.

In order to ensure that the crude assays HYSYS simulations are valid and can be used to study the effects of operating parameters towards the stabilized crude specifications and quality, the data generated from the HYSYS are compared with the plant data.

Figure 13 shows a graph of component mole fraction (dry basis) vs. crude components. The two different trends represent two different data which are the plant data from Terengganu Crude Oil Terminal, and also data generated from HYSYS (Crude Assay). The complete data in table form can be referred to in Appendix IV.

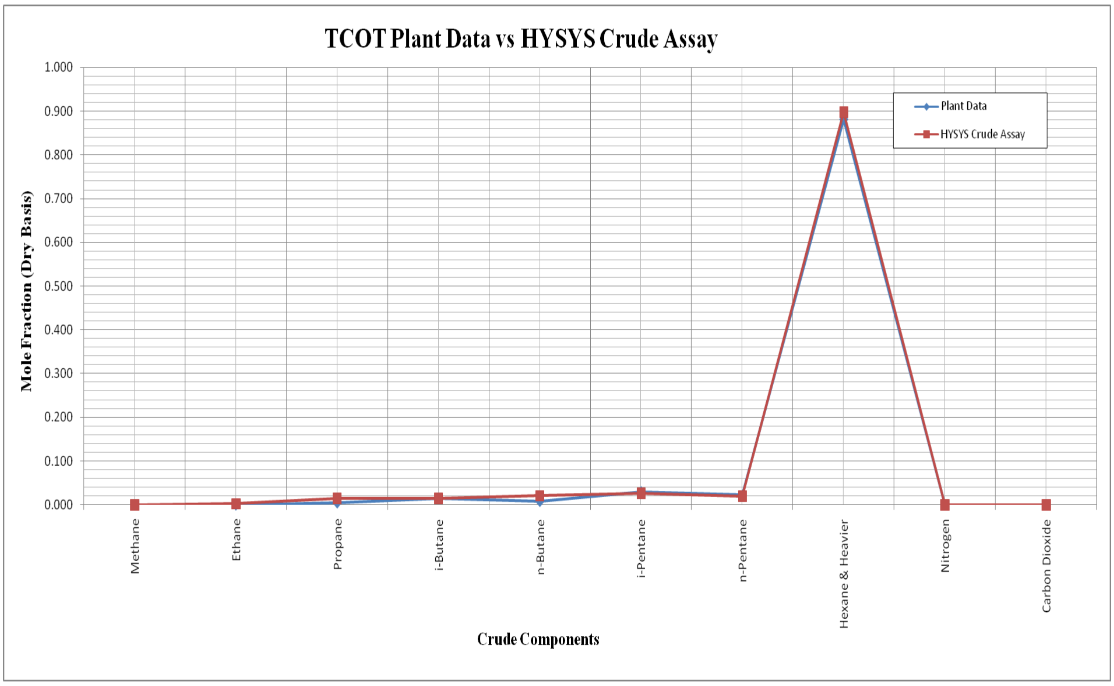

Figure 13: Comparison of Product Composition

For the purpose of simulation validation among all data trends, the validation is more focusing on two parts of the graph in which are the light components and also heavy hydrocarbons (C6+ and heavier) of the stabilized crude. In the first part of the graph (light components) from Methane to Pentane, it can be seen that two sets of data show the exact trends of component compositions. There is no clear fluctuation from the graph. As for the second part of the graph (heavy components ; C6 and heavier), it is clear that the data from HYSYS (Pseudo components & Crude assay), both are giving the same heavy hydrocarbon composition as per TCOT Plant data. The heavy crude is estimated to have a mole fraction in a range of 0.88 — 0.90 from total composition.

This is good as less volatile component are presents in the final products thus results in lower True Vapor Pressure which in turn is easier to store or export. The storage/export specification is set at 12 psia @ 38 °C. Any crude with pressure and temperature higher than specification was consider off-spec crude and should be removed from the vessel back to the crude stabilization trains.

On the other hand, for the non-condensable mixtures (Nitrogen and Carbon dioxide), the composition is nearly zero in which there is only a small trace of gas contaminants in the stabilized crude. From the above analysis, it can be concluded that all three data from HYSYS (ver. 2006) and TCOT plant data give out almost the same trends for all components. Thus, in general, the trends of the mole fraction of the components are similar for all sets of data. There are no major differences and it is proven that the simulation done using the HYSYS software is valid and can be used to simulate the real life plant process.

4.3.4 Total Validation (HYSYS Composition Data vs. Plant Data)

Total validation was done between HYSYS simulation (Pseudo & Crude Assay) vs Plant Data (stabilized crude composition) as per figure 14 below. The validation was carried out in order to identify any major differences between both simulations and compare it with the plant data. The composition of the crude are arranged according to the carbon number and heavy crude are lumped into one major component (hexane and heavier) which comprises crude from C6 up to C36+.

In general, the graph in figure 14 shows no major differences between each other, the mol fraction for each of the crude are exactly the same for both simulation which indicate both simulation method are reliable in predicting the crude oil behavior in crude stabilization plant. In addition, there is no major distinction between HYSYS Simulation software and the plant data as most of the composition data from HYSYS follows Plant Data accordingly.

For the first part of the graph (light hydrocarbon, C1 — C4), the composition of the HYSYS Simulation follows the trend of the Plant Data accordingly. On the other hand, heavy hydrocarbon which are lumped into one single component shows the highest mol composition about 0.88 mol thus make up most of the bulk crude compositions.

This shows that most of the volatile components have been flashed off during the crude oil stabilization by staged separation which consists of three Oil, Gas and Water separator vessel. High pressure differences between the live crude and the separator are the main driving force for the flashing off of the off-gas. The low pressure inside the separator cannot hold the crude in its liquid form thus, some of the volatile component turned into gas phase and separated from the heavy crude.

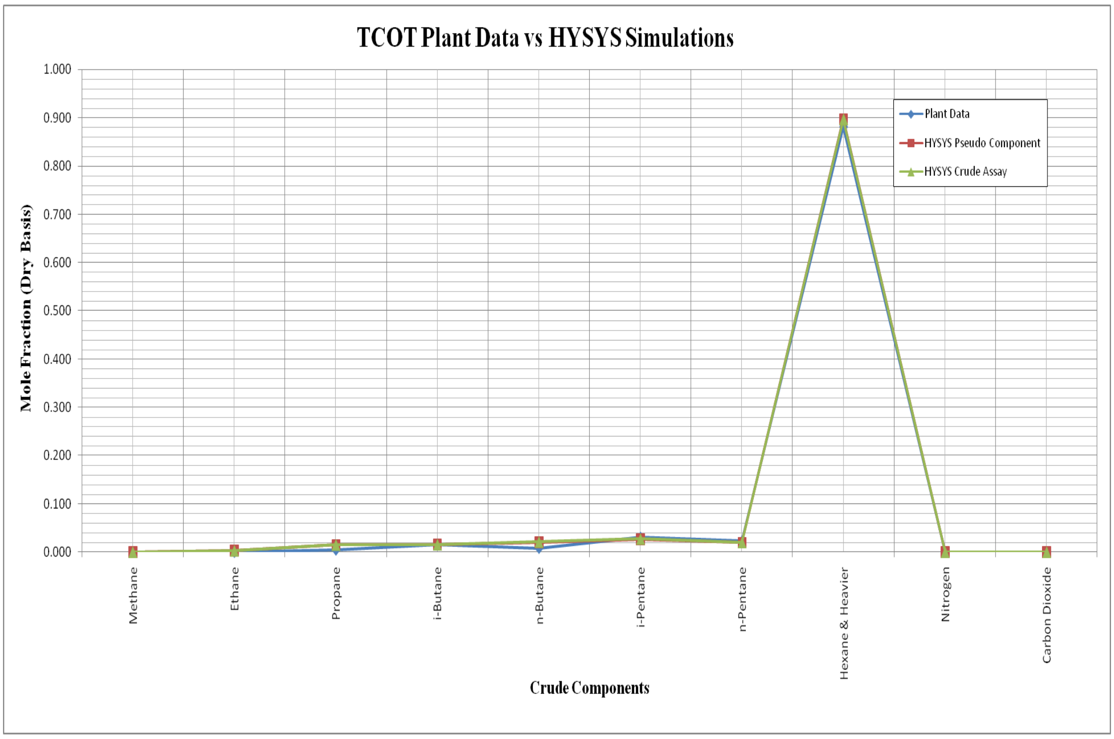

Figure 14 : Total validation of HYSYS Software (Version 7.3) vs. Plant Data

All in all, stabilized crude data generated by HYSYS simulation software are comparable to the plant data and valid. Stabilized crude composition contains about 90% of heavy hydrocarbon and less amount of light hydrocarbon which is a good factors for crude storage. Volatile composition such as methane and ethane was flashed off during stabilization process leaving only heavy hydrocarbon with lower vapor pressure. Normally, stabilized crude has true vapor pressure (TVP) of 12 psia @ 37.8 °C and stored in floating roof tank in atmospheric conditions.

4.4 PRELIMINARY RESULT ANALYSIS

4.4.1 HYSYS Simulations (Pseudo Components)

The inlet composition of the HYSYS simulation is compared with the final products (Stabilized crude) data generated. Figure 15 shows the trend of inlet and outlet data for HYSYS simulation (Pseudo Components) with free water content:

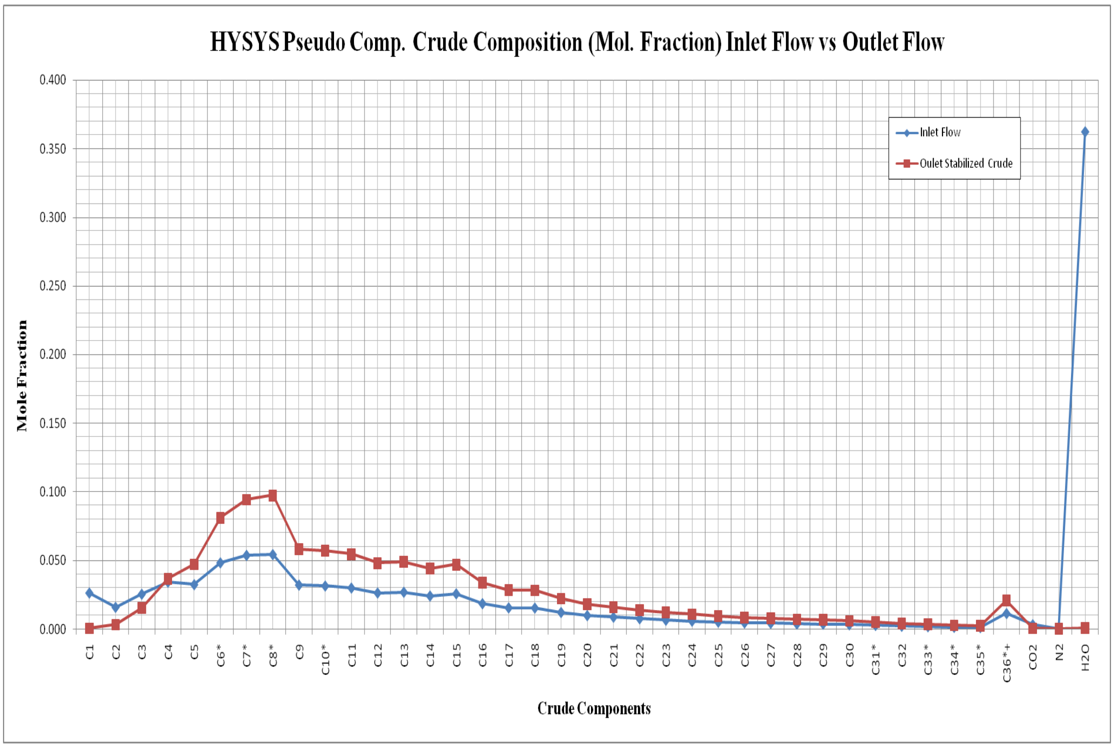

Figure 15: Inlet and Outlet Comparisons for HYSYS Simulation (Pseudo Components)

Based on the preliminary data analysis, it can be seen that from the graph that the water content in incoming flow is high at 0.35 of mole fraction. HYSYS simulations are able to remove almost 99% of the water content in the crude stabilization system thus reducing the water content in the stabilized crude less than 0.02. The inlet water volumetric flowrate to crude stabilization system is taken initially at 10% of the total dry crude inlet feed which accounts for 9.1% BS&W.

In addition, from Figure 15, it is observed that the stabilized crude from HYSYS contains more intermediate component compared to the inlet flow. This is mainly due to the high pressure of the three phase separator which traps the intermediate components in liquid phase while removing the volatile components as off-gas.

Further analysis also conducted as to analyze the distribution of the component in dry basis without considering water composition. Figure 16 shows the distribution of component in inlet and outlet crude of HYSYS simulations.

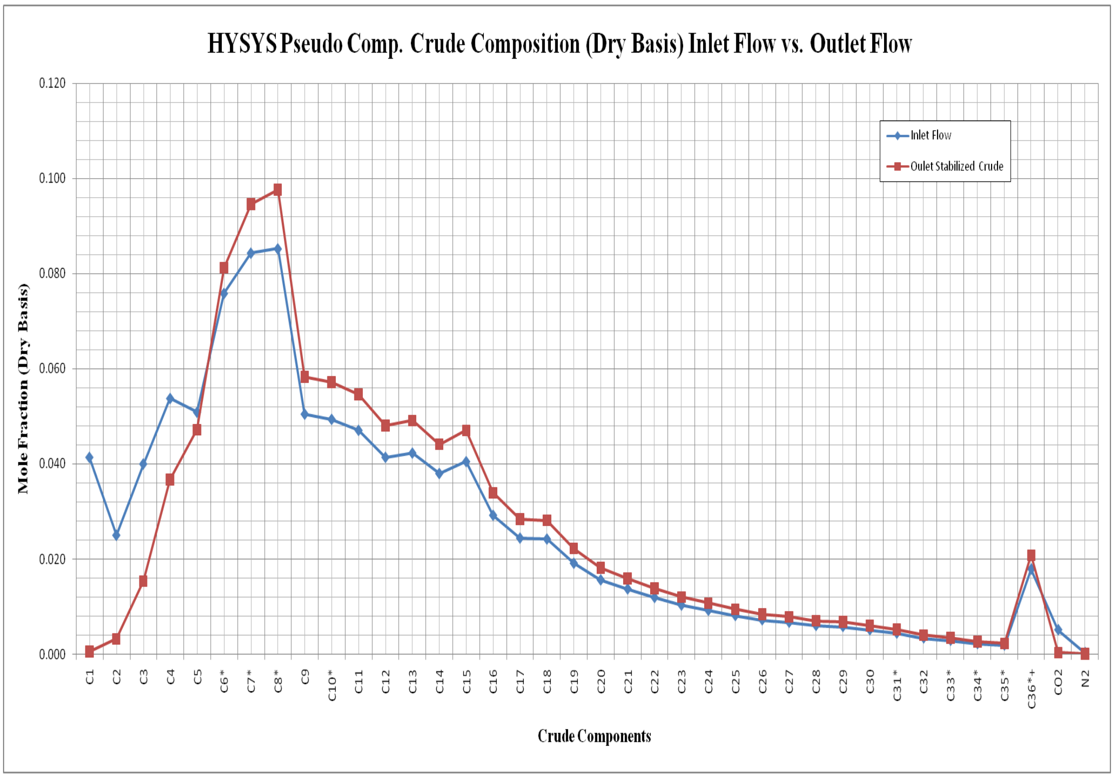

Figure 16: Inlet vs. Outlet Composition of HYSYS Simulations (Pseudo Components) Dry Basis

Based on the preliminary analysis of the distribution of components (in dry basis), it is observed that the inlet crude (live crude) contain high amount volatile components (C1 — C4) compared to the stabilized crude. This is mainly due to most of the volatile components are being flashed off during the crude stabilization process which result in high quality crude with standard specification of 12 psia for easy transportation and storage. The flashed off-gas from crude stabilization system will then be sent to the gas compression for further treatment and condensate recovery. The stabilized crude with TVP 12 psia, at 40 ^{o}C will then be transferred to floating roof crude storage tank at atmospheric condition (500, 040 barrel capacity)

In addition, the stabilized crude generated from HYSYS also shows high composition of intermediate - heavy hydrocarbons in the stabilized which results in high quality crude with high API Gravity. Based on the data obtained from TCOT, the stabilized crude is estimated to have a API gravity of 44.1 ^{o}. Thus, it can be concluded that the HYSYS simulation achieved its purpose to simulate a crude stabilization system.

Table 18 shows the properties of the stabilized crude obtained from the HYSYS simulation (Pseudo Components):

Table 18: Stabilized Crude Properties (HYSYS Pseudo Components)

Properties	Inlet Flow	Outlet Flow
Molecular Weight	65.5	180.2
Mass Density, kg/m3	792.2	755.4
Pressure, kPa	1801	83
Temperature, ^{0}C	26.96	37.78
Total Mass Flow, kg/h	1102162.6956	938318.4426
Volumetric Flowrate, L/hr	1420951.3816	1221920.536
Water Flowrate, Kbd	19.5000	0.010806462
Water Content, mg/L	90726.00209	58.46772572
BS&W, % Vol	9.0909%	0.0059%
Total Production, Kbd	215	184

Terengganu Crude Oil Terminal (TCOT) is designed to receive 450, 000 barrels of crude oil per day from the offshore platforms. Based on the site information from year 2011 — 2012, the average crude production was 195, 000 barrels per day. On the other hand, free water coming along with the dry feed is calculated at 10% of the dry feed inlet which is about 9.09% BS&W.

4.4.2 HYSYS Simulations (Crude Assays)

The inlet composition of the HYSYS simulation are compared with the final products (Stabilized crude) data generated. Figure 17 shows the trend of inlet and outlet data for HYSYS simulation (Crude Assays).

Based on the preliminary data analysis, it can be seen that from Figure 15 that the water content in incoming flow is as high at 0.337 of mole fraction. HYSYS simulations are able to remove almost 99% of the water content in the crude stabilization system thus reducing the water content in the stabilized crude less than 0.02. The inlet water volumetric flowrate to crude stabilization system is taken initially at 10% of the total dry crude inlet feed. Basically the trends shows by this graph is almost similar with HYSYS (Pseudo Components), in which there is no major differences between these two simulation. Thus, both simulation can be used to study the effects of operating parameters towards the crude oil stabilization operation.

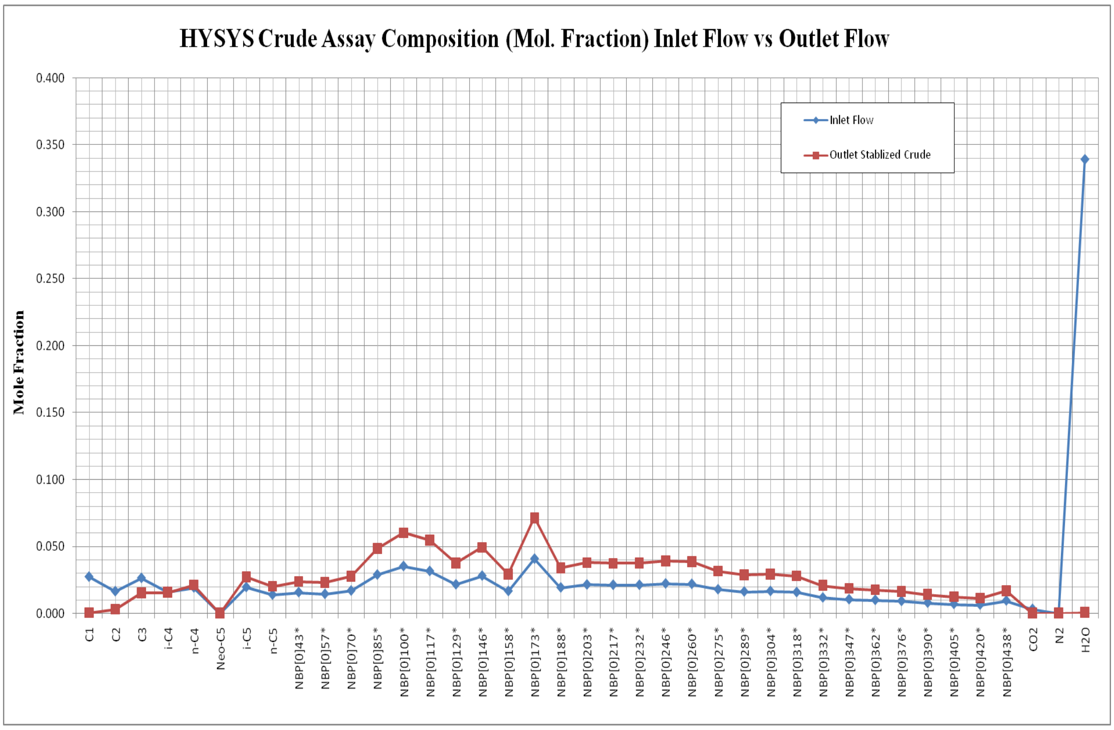

Figure 17: HYSYS Crude Assays Compositions (Inlet vs. Outlet)

In addition, from Figure 17, it is observed that the stabilized crude from HYSYS contains more intermediate component compared to the inlet flow. This is mainly due to the high pressure of the three phase separator which traps the intermediate components in liquid phase.

Further analysis also conducted as to analyze the distribution of the component on dry basis without considering water composition. Figure 18 below shows the distribution of component in inlet and outlet crude of HYSYS simulations.

Based on the preliminary analysis of the distribution of components (in dry basis), it is observed that the inlet crude (live crude) contain high volatile components compared to the stabilized crude. This is mainly due to most of the volatile components are being flashed off during the crude stabilization process which result in high quality crude with standard specification of 12 psia for easy transportation and storage.

The flashed light component in gas phase (during crude stabilization process) will then be sent to gas compression system for further treatment and condensate recovery. The light component (gas phase) is a valuable product either as a sales gas, fuel gas and also can be turned to liquefied petroleum gas products

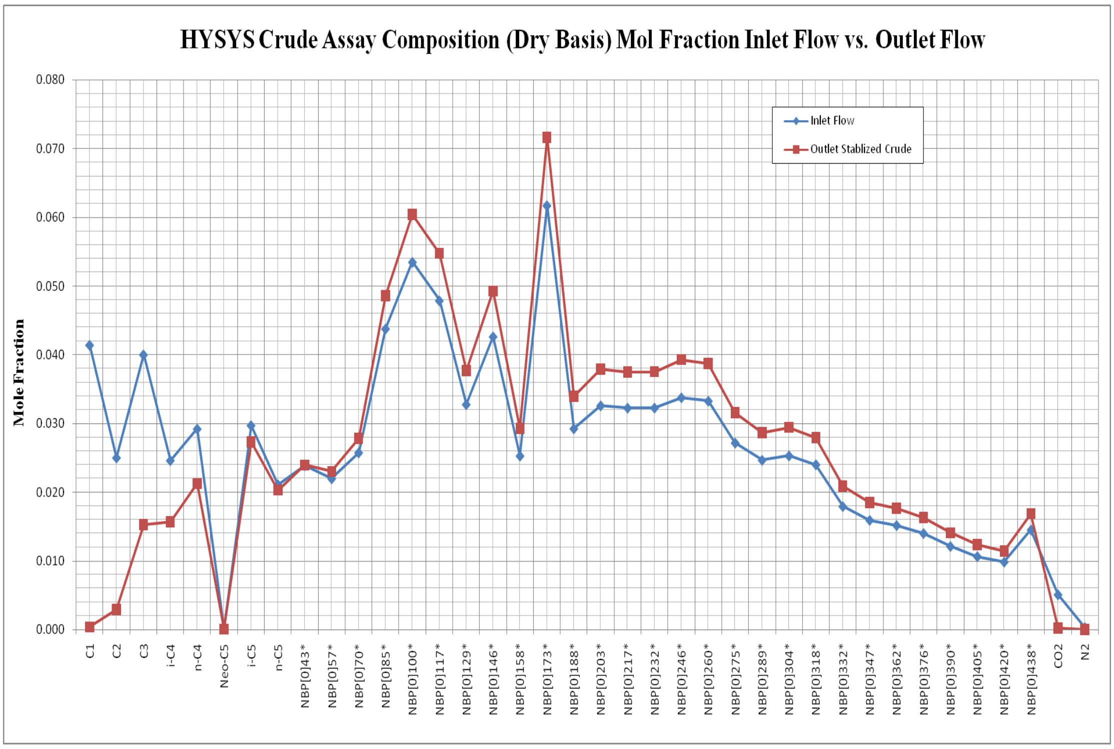

Figure 18: HYSYS Crude Assays (Inlet vs. Outlet) in Dry Basis Comparison

In addition, the stabilized crude generated from HYSYS also shows high composition of heavy hydrocarbons in the stabilized which results in high quality crude with high API Gravity. Based on the data obtained from TCOT, the stabilized crude is estimated to have a API gravity of 42 ^{o}. Thus, it can be concluded that the HYSYS simulation achieved its purpose to simulate a crude stabilization system.

Table 19 shows the properties of the stabilized crude obtained from the HYSYS simulation (Crude Assays):

Table 19: Stabilized Crude Properties (HYSYS Crude Assays)

Properties	**Inlet Flow**	**Outlet Flow**
Molecular Weight	67.77	176.9
Mass Density, kg/m3	839	808.9
Pressure, kPa	1801	83
Temperature, ^{0}C	26.95	37.78
Total Mass Flow, kg/h	1188171.9003	1019206.5154
Volumetric Flowrate, L/hr	1420951.3816	1213179.811
Water Flowrate, Kbd	19.50	0.01
Water Content, mg/L	90726.00209	66.0077852
BS&W, % Vol	9.0909%	0.0066%
Total Production, Kbd	215	183

4.4.3 HYSYS Methods Comparisons (Pseudo Comp. vs. Crude Assays)

Generally, there are two different methods to simulate crude oil stabilization system using HYSYS simulations which are by;

1. **Pseudo Component**
   Hypothetical components which represents heavy component of crude oil in the process simulations are created based on the laboratories characterization and crude sampling.
2. **Crude Assay**
   The petroleum characterization method in Aspen HYSYS converts laboratory analysis of crude oils, petroleum cuts and etc. into a series of discrete hypothetical components based on their boiling point.

However, there are several differences between both simulations which lead to variation of the results. Described below are the major differences of the two methods;

- **Molecular Weight**
  HYSYS Pseudo components have lower molecular weight (g/mol) compared to HYSYS Crude assays. The characterization of specific hypothetical components of pseudo components are set by the users based on laboratory analysis while the crude assays characterization are only based on light components properties.
- **Mass Density**
  HYSYS Crude assays have higher mass density (kg/m^3) compared to HYSYS Pseudo components. The mass density of each components (Crude Assays) are determined based on the fluid package of the simulations (Peng –Robinson) while mass density of the pseudo component are set by the users.

All in all, HYSYS Pseudo components are more of simulation based on laboratory results (experimental based simulation). On the other hand, HYSYS Crude assay determined the properties of heavy components based on normal boiling points of each component.

4.5 EFFECTS OF DIFFERENT OPERATING CONDITIONS

In real life plant, the process is not at a steady state since there are always fluctuations in the process parameter ; operating conditions which usually depends on the operation mode. This may be due to many reasons such as changing surrounding conditions, upset in other related process unit upstream and breakdown of related equipments. As a result of these parameter changes, the specifications of the product may also change. Therefore, it is important to know how much of these changes that the process can tolerate and at which point the parameter change will cause the product to become off-specification. In order to obtain those data, a one dimensional study is done on the simulated crude oil stabilization plant by manipulating several parameter listed below:

1. **Inlet Feed Parameters**
   a. Dry feed volumetric flowrate
   b. Free water flowrate
   c. Inlet Temperature
   d. Inlet pressure
2. **Three Phase Separator Parameters**
   a. V-220 operating pressure
   b. IV-225 operating pressure
   c. V-230 operating pressure
3. **Pre-Heater Train Performance**
   a. Fuel gas inlet flowrate
   b. Furnace efficiencies
   c. Furnace duty

Manipulated variable as per above will be varied to study the effects of that particular variable on stabilized crude specifications. The product specifications that are being monitored in this study are listed as per below.

1. True Vapor Pressure, TVP : 12 psia
2. Basic Sediment & Water, BS&W content <0.5%

In order to study the effects of the manipulated parameters, all other values except the parameter being studied need to be kept constant. Table 20 shows what parameter is kept constant for each study where C represents constant and V represents variable. The findings from the observations will be discussed further in the following section. The complete data obtained in the study can be referred in Appendix VI : Effects of Different Operating Conditions.

Manipulated Parameter		Study the Effect of								
		Dry Feed Flowrate	Water Flowrate	Feed Temp	Feed Pressure	Separator Pressure	Separator Temp.	HX-210 Outlet Temp.	HX-220 Outlet Temp	Furnace Efficiencies
Feed Properties	Dry Feed Flowrate	V	C	C	C	C	C	C	C	C
	Water Flowrate	C	V	C	C	C	C	C	C	C
	Temperature		C	V	C	C		C	C	C
	Pressure	C	C	C	V	C	C	C	C	C
Separator	Pressure	C	C	C	C	V	C	C	C	C
	Temperature	C	C	C	C	C	V	C	C	C
Pre-Heater Trains	HX-210 Outlet Temp.	C	C	C	C	C	C	V	C	C
	HX-220 Outlet Temp.	C	C	C	C	C	C	C	V	C
	Furnace Efficiencies	C	C	C	C	C	C	C	C	V
Pump, P-210 Power		C	C	C	C	C	C	C	C	C
HX-210 Duty		C	C	C	C	C	C	C	C	C
HX-220 Duty		C	C	C	C	C	C	C	C	C

Table 20 : Status of Operating Conditions for the Study of Effects of Changing Operating Conditions

4.5.1 True Vapor Pressure, TVP

The main product specifications that is considered for a crude stabilization plant is the True Vapor Pressure, TVP of the stabilized crude. Therefore, the TVP of the product is the most important specification that needs to be monitored closely during the operation of the crude stabilization plant. The lower the TVP of the product, the higher the quality is. The standard method for measuring TVP is ASTM 2879. As stated earlier, in this project, the effect of parameter changes on the TVP has been studied by changing all manipulated parameters, i.e . inlet feed properties, three phase separator & hot oil system. Any significant impacts of the variables on the operation will be observed.

4.6 EFEECTS OF INCOMING CRUDE OIL VARIABLES

In the first part of these simulations, the inlet crude properties are varied to study the effects of the inlet parameters towards crude oil stabilization operations. Normal incoming crude inlet to the terminal is at 195 Kbd at 27 °C, 17 Bar with BS & W of 9 vol%. Four major inlet properties which are flow rate, temperature, pressure and free water content is set as manipulated variable with True Vapor Pressure, TVP of the stabilized crude is the controlled variables.

4.6.1 Effect of Feed Flow Rate

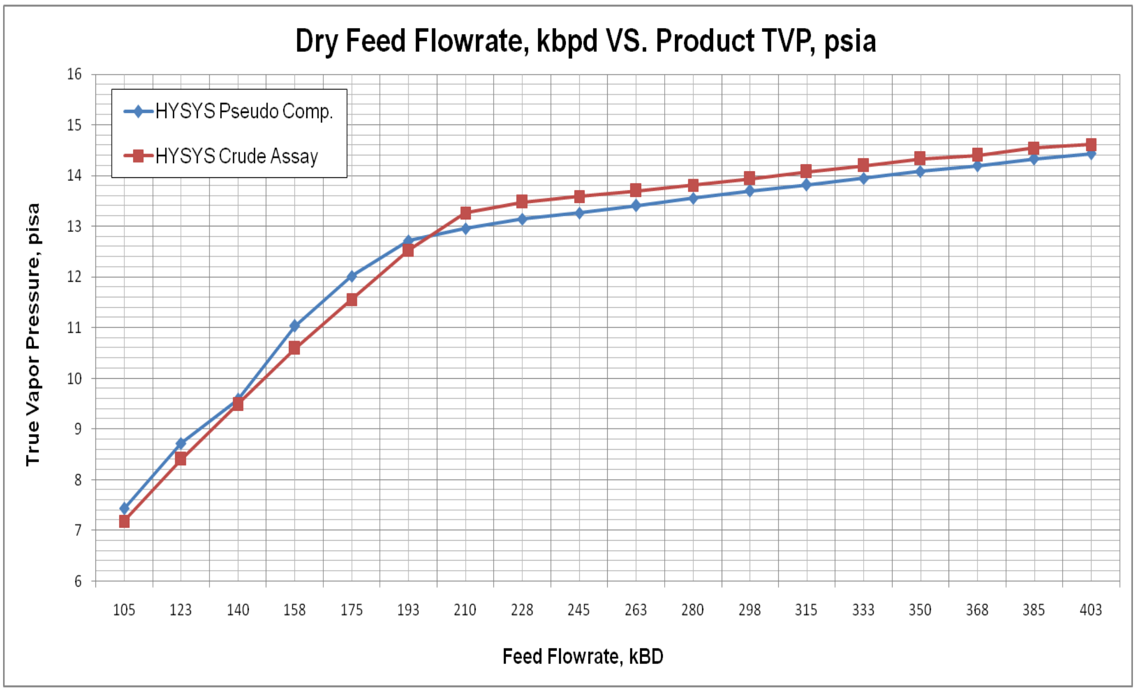

Figure 19 : Effect of Dry Feed Flow Rate towards Product TVP

The normal current feed flow rate used for the base case study is 195 kBD. The flow rate is then decreased to 50% and then increased to 230% in 10% intervals. Figure 19 shows how the change in feed flow rate affects the affects the TVP of the stabilized crude.

From the graphs, it can be seen for both HYSYS simulation methods, as the flow rate increases, the TVP also increases. This increase in TVP is because when the flow rate increases, more heat is required to flash off the light component in the dead crude. In order to simulate the effects of the feed flow rate, all other variables and unit operation such as heat exchangers duty (HX-210 & HX-220) are kept constant. This results in insufficient heat to flash off the entire volatile component with traces amount of the C1 — C4 still there in the stabilized crude thus increasing total mixture TVP.

Therefore, the TVP would gradually increases with the increase of feed flow rate. For an acceptable range of TVP from 10 psia — 12 psia, the maximum flow rate percentage that can be processed by the crude stabilization plant is at 139 —188 kBD respectively.

4.6.2 Effect of Feed Temperature

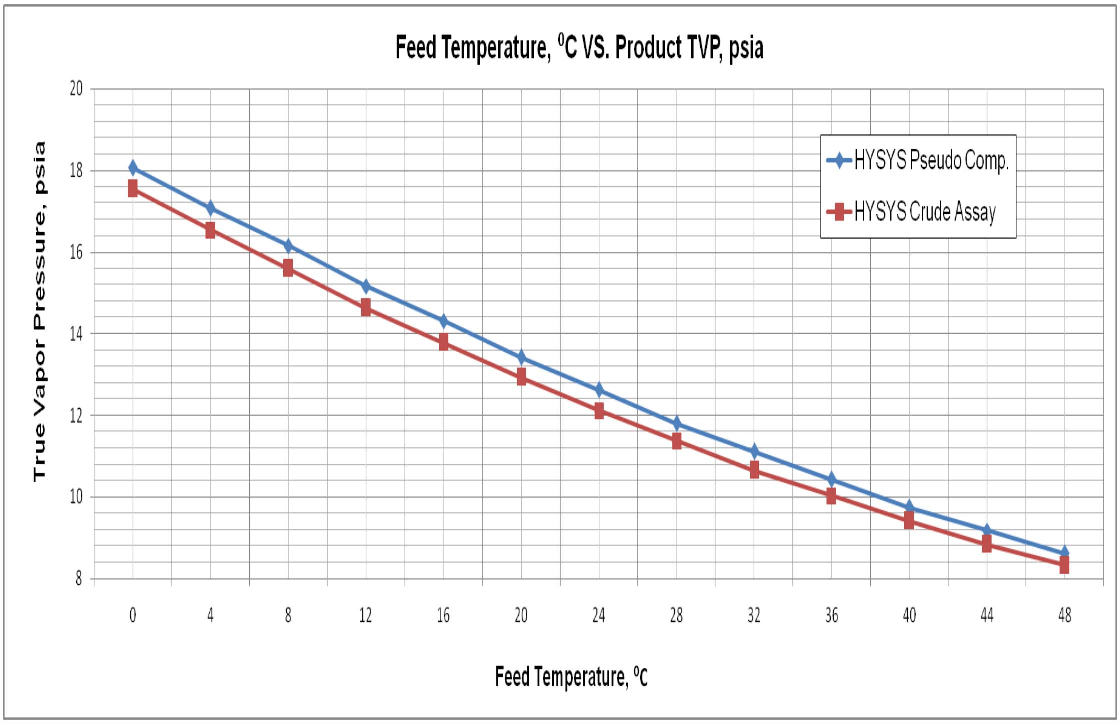

Figure 20 : Effect of Feed Temperature towards Product TVP

The inlet feed to the crude stabilization plant is normally at 27 °C from the offshore platforms. In order to study the effects of feed temperature towards product TVP, the temperature is decreased to 2 ^oC and then increased to 40 ^oC at 4 ^oC intervals. Figure 20 shows how the change in feed temperature affects the TVP of the stabilized crude.

As can be seen in the graphs, as the temperature of the feed is increased, the product TVP gradually decreased. The increase in the feed temperature would cause more portions of the light component to flash off easily from the crude and thus reduce the TVP of the product.

The minimum temperature that the crude stabilization plant can tolerate n order to achieve the specified TVP (10 — 12 psia) is in a range of 26 ^oC to 38 ^oC. Any temperature lower than 26 ^oC would cause the stabilized crude become off-specifications as it is will require a larger duty for the heat exchanger to heat the crude before entering the pressure vessels.

4.6.3 Effects of Feed Pressure

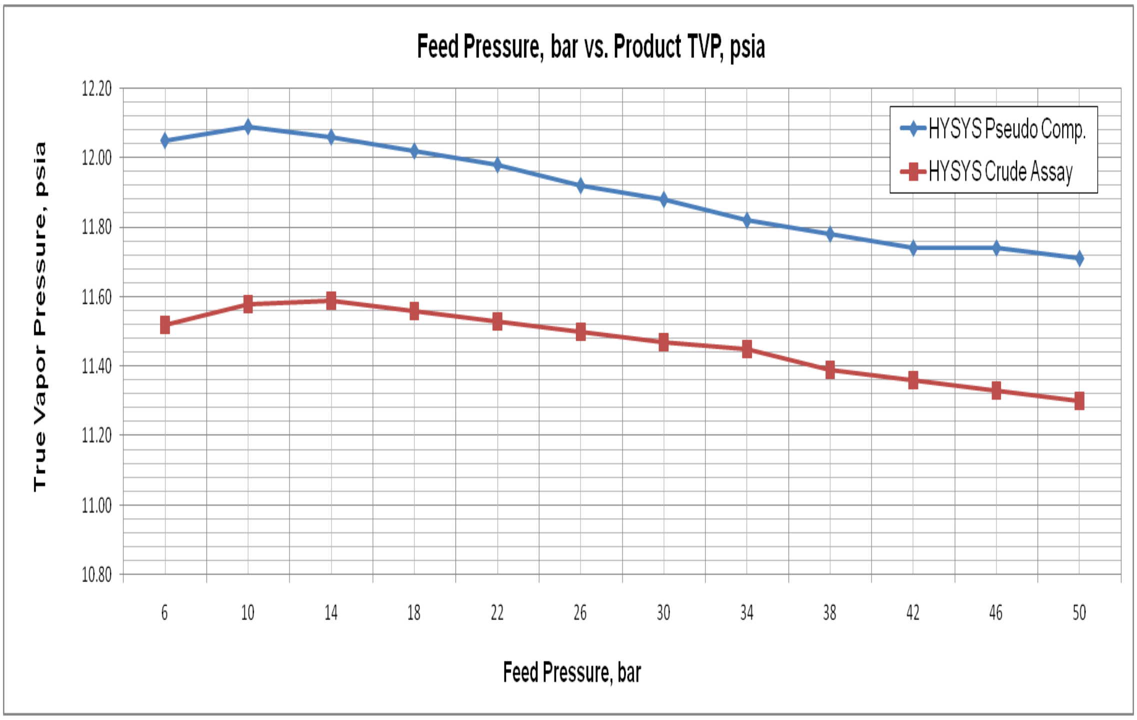

Figure 21 : Effects of Feed Pressure towards Product TVP

At normal operating conditions, the pressure of the feed to the crude stabilization plant is 18 bar. At first, to study the effects of the feed pressure towards the product TVP, the feed pressure is reduced to 6 bar and then increased to 50 bar at 4 bar intervals. Figure 21 shows how the change in feed pressure affects the TVP of the stabilized crude.

Form the graph, for both simulations; it is observed that the highest TVP recorded (12.1 psia) is at 10 bar feed pressure. As the pressure of feed increased, the TVP of the stabilized crude gradually decreasing. This shows that the impact of the feed pressure towards the crude stabilization unit is insignificant. The product TVP decrease as the feed pressure increase is due to the high pressure drop into the pressure vessels which lead to high amount of volatile component being flashed off to the stabilization gas header. Thus, the stabilized crude contained only traces of light component which posses lower TVP and can be stored in a atmospheric condition safely.

To summarize, even though the change in the feed pressure will affect the product TVP, the TVP will not go off-specification since the highest TVP recorded is only at 12.45 psia. This shows that the impact of incoming crude feed is insignificant towards the crude oil stabilization operations.

4.6.4 Effect of Water Flow Rate

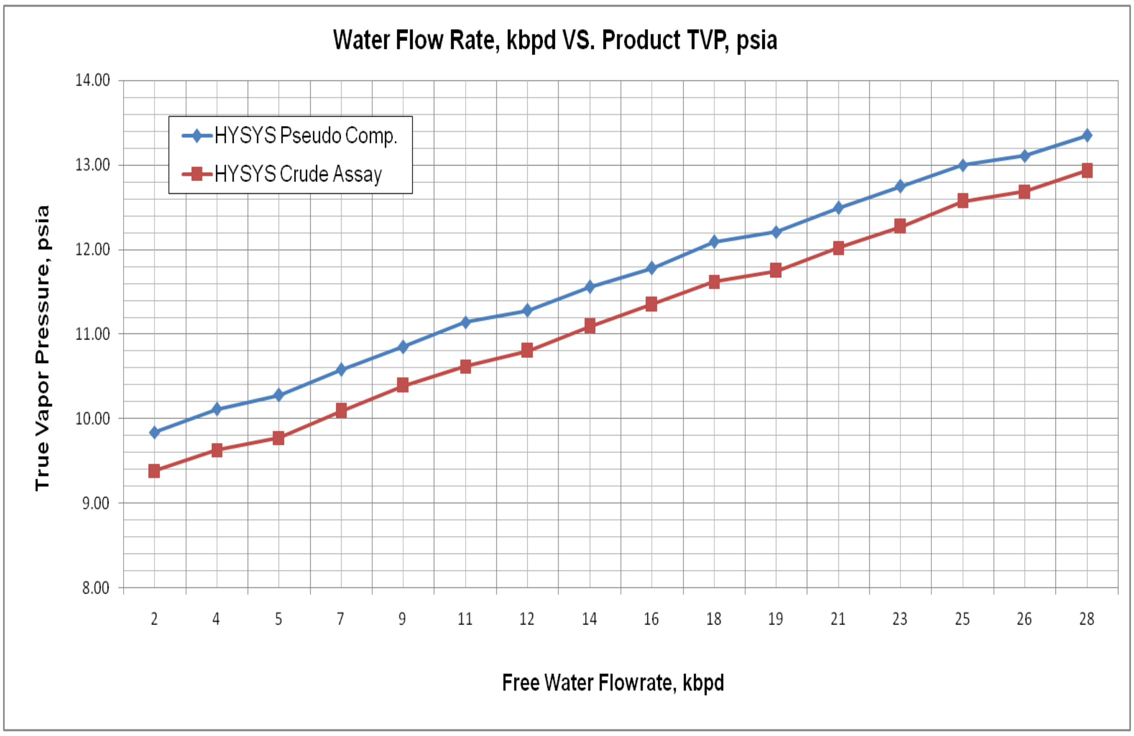

Figure 22 : Effects of Water Flow Rate towards Product TVP

Based on the Terengganu Crude Oil Terminal, TCOT Design Basis Memorandum, the existing facility is capable of processing 21, 600 barrels per day of free water content in the inlet crude in which equivalent to 4.8 vol% for 450, 000 barrels of crude oil production per day.

However, based on site information from year 2011 - 2012, average crude oil production is around 195 kBD. Thus the free water content is assumed 10% of the dry feed which is about 19.5 kBD (9.09 vol% BS&W). In order to study the effects of water inlet flowrate towards the crude stabilization operation, water flow rate is decreased to 2 kBD and increased to a maximum of 29 kBD. Figure 22 shows the effects of the water inlet flow rate towards the product TVP.

Based on the graph for both HYSYS simulation methods, as the water content inside the crude oil increase, the resultant stabilized crude will also have higher TVP. The increase of water content will increase total mixture density and mass flow which require a high duty of heaters/ heat exchangers to heat the process fluid to a suitable temperature before entering the pressure vessel. In case of optimum operating temperature cannot be achieved, it will affect the separation inside the pressure vessel thus results in less volatile component being flashed off to the stabilization gas header.

The maximum water content that the crude stabilization operation can tolerate in order to achieve the specified TVP is 20 kBD for a dry feed flow rate of 195 kBD which is estimated around 10 vo1% BS&W. Any value higher than recommended will results in off-specification crude (TVP high than 12 psia)

4.7 EFFECTS OF OIL, GAS & WATER SEPARATORS VARIABLE

Basically, the purpose of the simulation is to study the effect of changing the operating condition of pressure vessel towards the crude oil stabilization operation. There are in total of three pressure vessel at Terengganu Crude Oil Terminal,TCOT which are the High Pressure Separator (V-220 A/B), Electrostatic Precipitator (V-225 A/B), Low Pressure Separator (V-230 A/B).

4.7.1 Effects of High Pressure Separator Operating Pressure

High Pressure Separator (V-220 A/B) is the first three phase separator in the crude oil stabilization system which normally operating at 466 kPa at 75 – 85 °C. The light components flashed off from the high pressure separator are sent to the gas stabilization header prior for condensation and purification process.

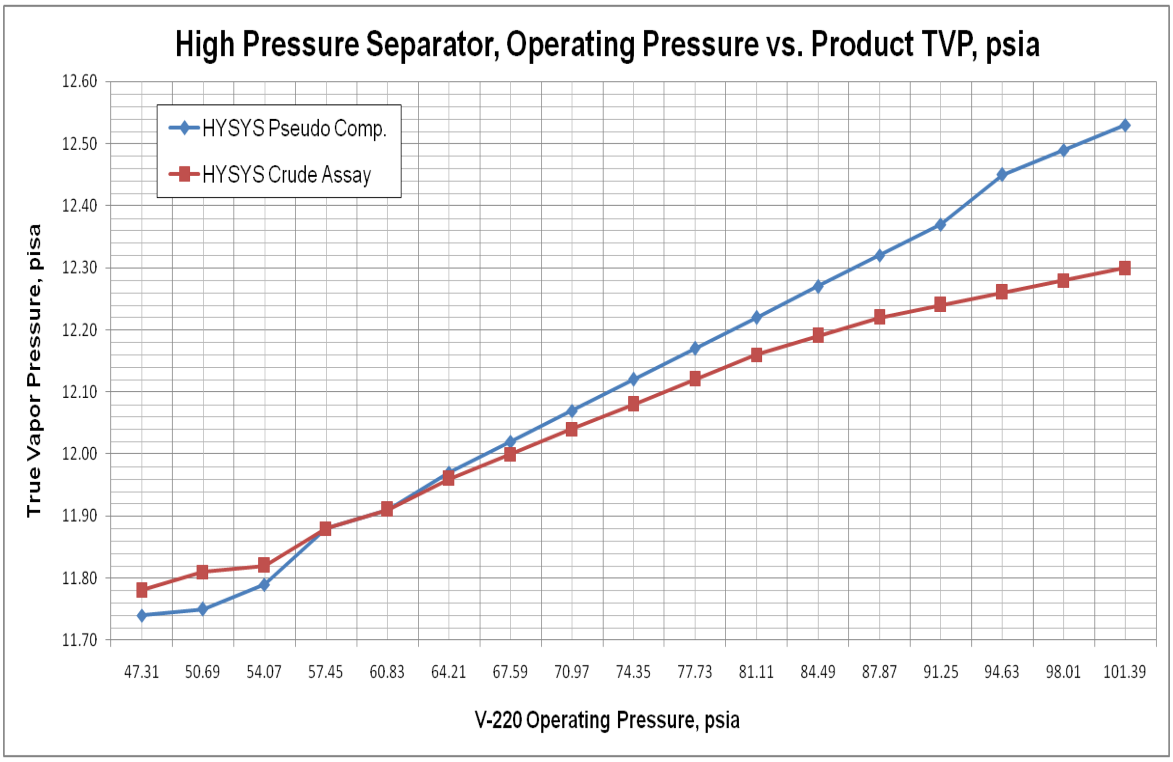

Figure 23 : Effects of High Pressure Separator,V-220 A/B Operating Pressure

Based on site visit to Terengganu Crude Oil Terminal on 14th May 2012, V-220 A/B is operating at 466 kPa. For the purpose of these simulations, the operating pressure of V-220 A/B is decreased from 330 kPa and increased up to 703 kPa.

The graph in Figure 23 shows that as the operating pressure of V-220 A/B is increased, the stabilized crude product TVP also increased. The increase in TVP of the products is because the increment in V-220 A/B operating pressure means lowers differential pressure between the incoming crude inlet and the pressure vessel. This results in fewer amounts of volatile components being flashed off as gas phase at the high pressure separator. Thus, there are still traces amount of volatile component in the rundown crude to storage which contributed to increase of product TVP.

However, the impact is not very significant as the increment in products TVP happens slowly and not very drastically. In order to ensure smooth operation without any upset to the operation, the High Pressure Separator,V-220 A/B operating pressure should be kept around 450 kPa – 500 kPa.

4.7.2 Effects of Electrostatic Precipitator Operating Pressure

Electrostatic Precipitator, V-225 A/B is two phase separator which uses electrostatic force to break the oil-water emulsion formed inside the crude. It is normally operating at 300 kPa at 75 – 85 °C. Separated crude oil and off-gas from V-225 A/B are gathered together and sent to the Low Pressure Separator, V-230 A/B.

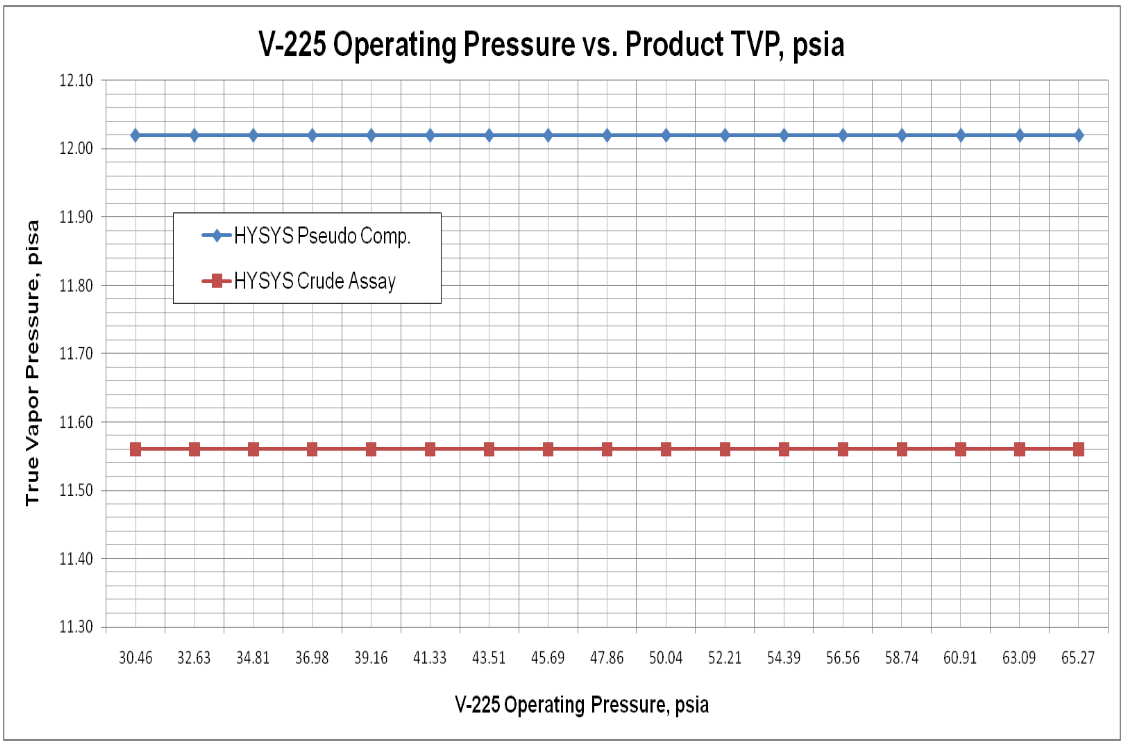

Figure 24 : Effects of Electrostatic Precipitator Operating Pressure

Based on site visit to Terengganu Crude Oil Terminal on 14th May 2012, V-225 A/B is operating at 300 kPa. For the purpose of these simulations, the operating pressure of V-225 A/B is decreased from 213 kPa and increased up to 448 kPa.

The graph in Figure 24 shows that as the operating pressure of V-225 A/B is increased, the product TVP remain constant. This is because the off-gas separated from V-225 A/B is sent to Low Operating Pressure, V-230 A/B which means no volatile component are removed from the crude.

The impact of V-225 A/B operating pressure is insignificant to the operation as the increment in the operating pressure only results in constant products TVP. In order to ensure smooth operation without any upset to the operation, the High Pressure Separator,V-225 A/B operating pressure should be kept around 300 kPa – 350 kPa.

4.7.3 Effects of Low Pressure Separator Operating Pressure

Low Pressure Separator (V-230 A/B) is the last three phase separator in the crude oil stabilization system which normally operating at 179 kPa at 75 – 85 °C. The light components flashed off from the low pressure separator are sent to the gas stabilization header prior for condensation and purification process.

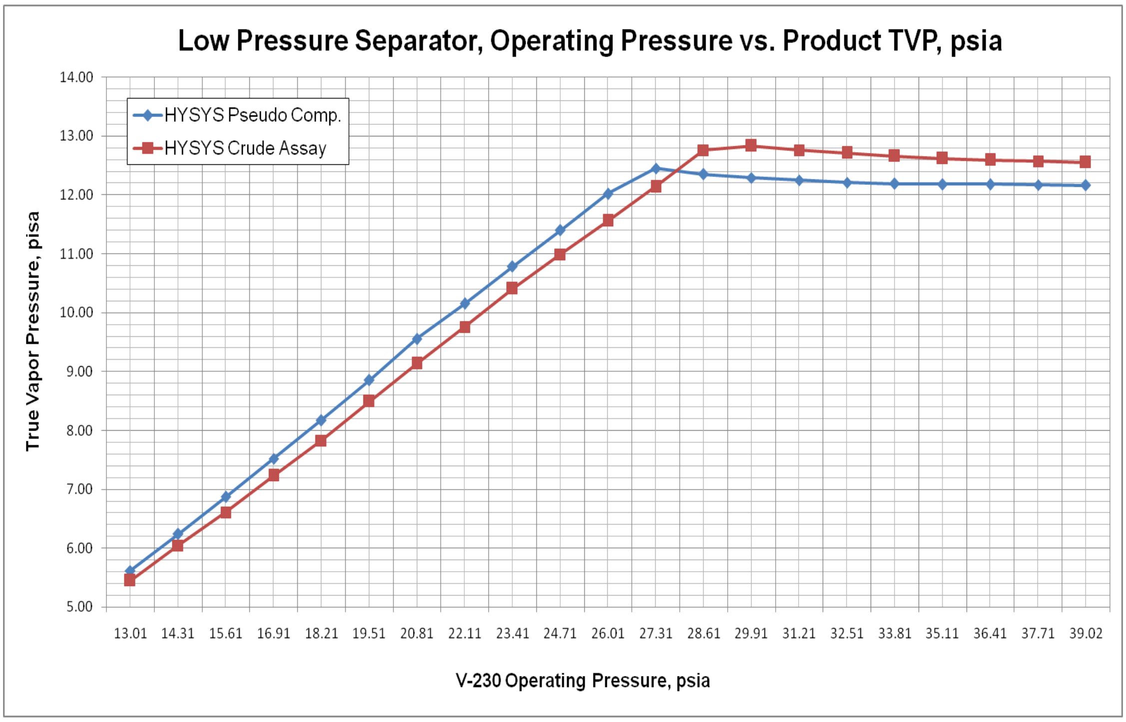

Figure 25 : Effects of Low Pressure Separator Operating Pressure

Based on site visit to Terengganu Crude Oil Terminal on 14th May 2012, V-220 A/B is operating at 179 kPa. For the purpose of these simulations, the operating pressure of V-220 A/B is decreased from 103 kPa and increased up to 268 kPa.

The graph in Figure 25 shows that as the operating pressure of V-230 A/B is increased, the stabilized crude product TVP also increased. The increase in TVP of the products is because the increment in V-230 A/B operating pressure means lowers differential pressure between the incoming crude inlet and the pressure vessel. This results in fewer amounts of volatile components being flashed off as gas phase at the high pressure separator. Thus, there are still traces amount of volatile component in the rundown crude to storage which contributed to increase of product TVP. In order to ensure smooth operation without any upset to the operation, the Low Pressure Separator, V-230 A/B operating pressure should be kept around 170 kPa – 230 kPa.

4.8 PRE - HEATER TRAINS VARIABLE

Pre-heater trains consist of two major heat exchanger banks (HX-210 & HX-220) which supply required heat to the live crude before entering the crude stabilization trains. There are eight Crude to crude Heat Exchanger, HX-210s per train, arranged in two parallel banks of four exchangers in series. The HX-210s supply needed heat to warm the inlet crude from 29.4 °C to 52 °C, and at the same time cools the hot stabilized crude to 40 °C before it is transferred to storage.

On the other hand, there are two parallel HX-220s per train ; one downstream of each parallel bank of four HX-210s. The HX-220s supply the additional heat to raise the inlet crude temperature to 80 °C using hot oil at 260 °C.

4.8.1 Effect of Crude to Crude Exchanger, HX-210s Outlet Temperature

In order to study the effect of changing HX-210s outlet temperature, all other parameters are kept constant including the next heat exchanger, HX-220s duty. Thus, HX-220s resultant temperatures are also observed along with the product TVP.

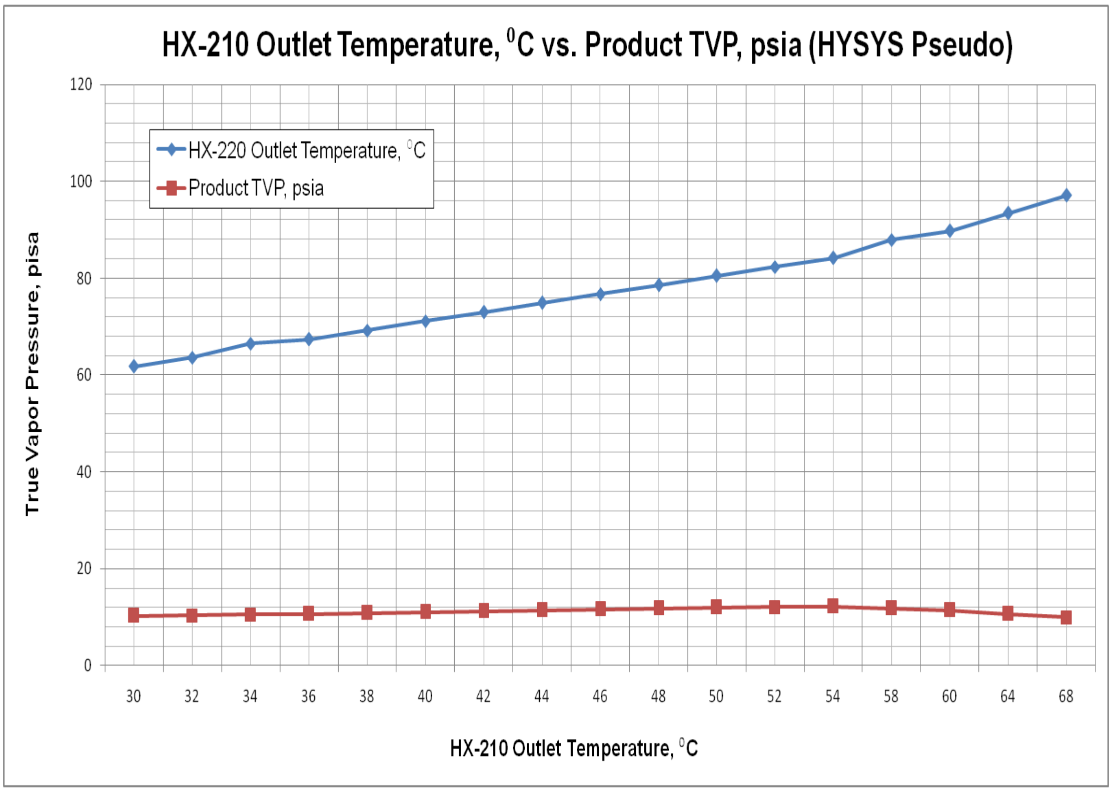

Figure 26: Effects of HX-210s Outlet Temperature (HYSYS Pseudo Components)

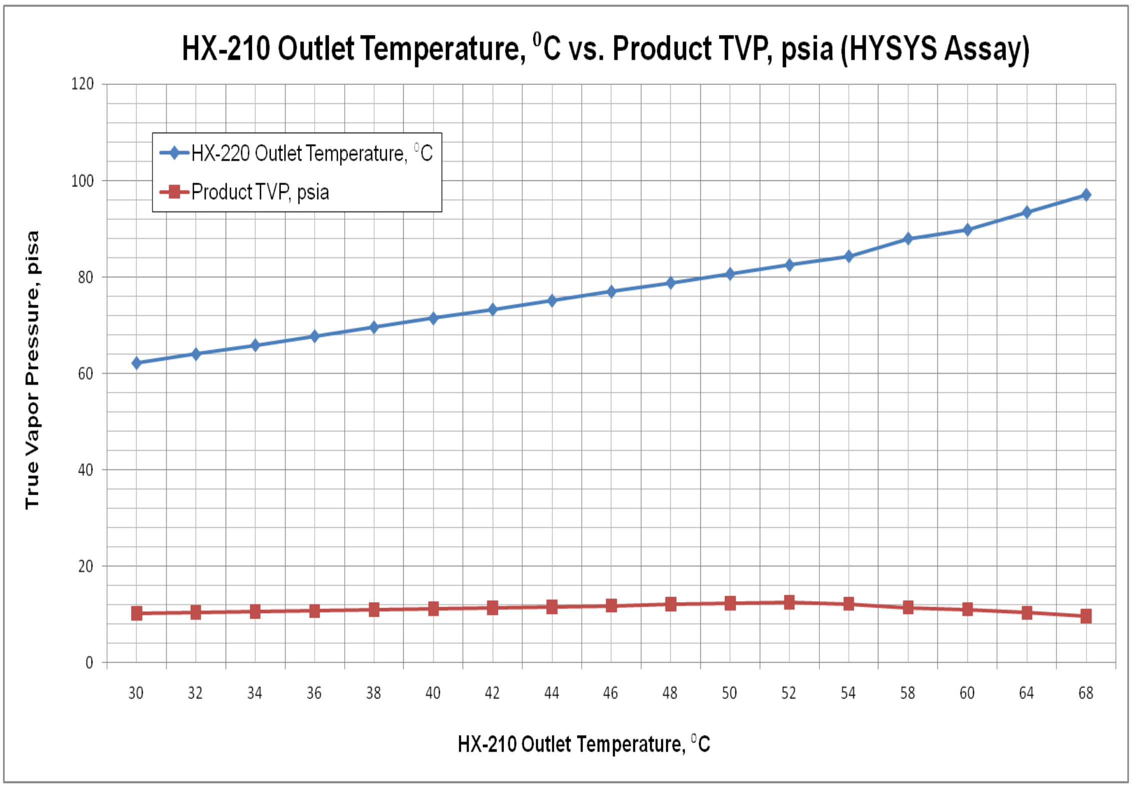

Figure 27: Effects of HX-210s Outlet Temperature (HYSYS Crude Assays)

Figure 26 & 27 show graphs for the effects of HX-210s outlet temperature for HYSYS simulation (both Pseudo Component and Crude Assays). There are two parts of the graph as a result of changing HX-210s outlet temperature, which are the product TVP and also HX-220s resultant temperature.

For the purpose of this study, the outlet temperature of HX-210s is decreased to 30 °C and increased up to 68 °C instead of normal operation at 52 °C. As HX-210s outlet temperature increase, the outlet temperature of subsequent HX-220s also increased. Product True Vapor Pressure also increased but then starting to decrease when the temperature of HX-210s is 52 °C (normal operation). Thus, it is observed that the effect of changing HX-210s outlet temperature is insignificant towards the crude oil stabilization system. In order to ensure smooth operation for the pre-heaters train, the outlet temperature of HX-210s should be kept more than 52 °C. Heat supplied is used to flash off the volatile component of the live crude.

4.8.2 Effect of Hot Oil to Crude Exchanger, HX-220 Outlet Temperature

In order to study the effect of changing HX-220s outlet temperature, the outlet temperature of HX-210s is kept constant at 52 °C. In normal operation, HX-220s outlet temperature is 80 °C.

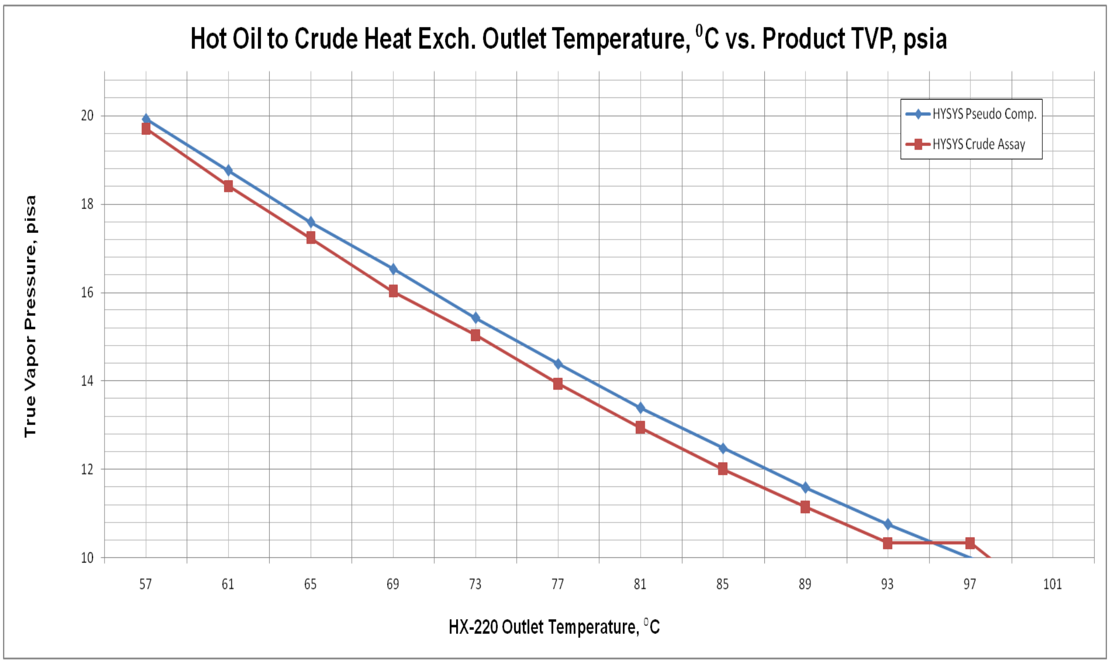

Figure 22 : Effects of HX-220s Outlet Temperature

Figure 28 shows graph for effects of HX-220s outlet temperature for HYSYS simulation (both Pseudo Component and Crude Assays) towards Product True Vapor Pressure. For the purpose of this study, the outlet temperature of HX-220s is decreased to 57 °C and increased up to 97 °C instead of normal operation at 85 °C. As HX-210s outlet temperature increases, the True Vapor Pressure of the stabilized crude decreases.

This is because more heat is supplied to flash off the light (volatile) components inside the live crude coming from the offshore. Thus, leaving only heavy and viscous components in rundown crude to storage with low vapor pressure. All in all, the impact of changing the outlet of HX-220s is very significant towards the crude oil stabilization operation. Thus, in order to ensure smooth operation for the pre-heaters train, the outlet temperature of HX-220s should be kept around 85 – 90 °C with HX-210s outlet temperature is constant at 52 °C. Hot oil and furnace performance will be discussed further to analyze the limitation of the pre-heaters train.

4.8.3 Fired Heater (HX-610) Analysis

4.8.3.1 Design Specifications and Current Operation

Terengganu Crude Oil Terminal, TCOT was designed with four (4) fired heaters (HX-610 A/B/C/D) to heat up hot oil (PETRONAS Danol XHT 32) from 150 to 260 °C with design efficiency of 89% for each fired heater. At full plant production load, 445 kbpd of crude flow, fired heater shall be operated based on N+2 (100% sparing) philosophy in which 12 burners (each fired heater with 6 burners) will be online.

Currently, three fired heaters (12 burners online) are in operation with production load of 175 kbpd of crude. The major differences between current operation and design specifications consist of the following:

Table 21 : Different Between Design and Current Operation

PARAMETER	DESIGN SPECIFICATION	CURRENT OPERATION
Crude Flow, kbpd	442	175
Hot Oil Inlet Temperature, °C	150	130 – 150
Hot Oil Outlet Temperature, °C	260	190 – 210
Stack/Flue Gas Temperature, °C	197	190 – 200
No. of Fired Heater In Operation	2+2 (12 burners)	3+0 (12-13 burners)

**Fired heater HX-610A is under maintenance*

4.8.3.2 Fired Heater Actual Efficiency

Efficiency calculation had been carried out to evaluate actual fired heater performance. Average value of fired heater HX-610C and HX-610D were used to estimate current performance of fired heater HX-610B, due to unavailability data for the calculation purposes. Assumptions were made on the physical properties of hot oil and fuel gas. The calculation was performed (Refer Appendix V) using the following formula and results are depicted in Table 22.

$$\boldsymbol{Heater\ Efficiency} = \frac{Heat\ Absorbed\ by\ Hot\ Oil\ (MW)}{Heat\ Released\ by\ Fuel\ Gas\ (MW)} \times 100\%$$

Table 22 : Calculated Combined Thermal Efficiency of Fired Heater HX-610C and D

Date	Hot Oil				Total Heat Released by Fuel Gas (MW)	Thermal Efficiency (%)
	HX-610B Heat Adsorbed (MW)	HX-610C Heat Adsorbed (MW)	HX-610D Heat Adsorbed (MW)	Total Heat Adsorbed (MW)		
July 2010	6.91	6.32	7.52	20.75	53.87	38.51
August 2010	6.59	6.86	6.33	19.78	51.72	38.25

Based on Table 22, it is observed that currently, the efficiency of the furnace system (HX-610s) has dropped below the design specification. The calculated thermal efficiencies are a combined value for HX-610B C and D since only total fuel gas flow is available. Average calculated efficiencies during year 2010 was 38.38% which is 56.88% deviation from the design efficiency (89%). This indicates the fired heater performance will not achieve 89% of efficiency even all the burners are put online. Further maintenance and regular monitoring are important to increase the efficiency of the pre-heater train's performance.

4.9 HOT OIL TO CRUDE EXCHANGER, HX-220X PERFORMANCE

4.9.1 Scenario 1 — Variable Water Volume % at Constant Outlet Temperature & Inlet Crude Flow

Table 23 : Scenario 1 for HYSYS Simulation Pseudo Component's Method

Scenario 1 (Pseudo Comp.)	1	2	3	4	5	% Deviation
HX-220 Inlet Temperature	57					
HX-220 Outlet Temperature	87					
Inlet Crude, kbpd	175					
Water (vol%)	4.8	7	10	15	20	317
Water, kbpd	8.8	13.2	19.4	30.9	43.8	317
HX-220 Duty, MW	19.60	20.65	22.13	24.88	27.97	43
V-220 High Pressure Vessel Gas Production, MMSCFD	12.71	12.74	12.79	12.85	12.91	2

Table 24 : Scenario 1 for HYSYS Simulation Crude Assay's Method

Scenario 1 (Crude Assays)	1	2	3	4	5	% Deviation
HX-220 Inlet Temperature	57					
HX-220 Outlet Temperature	87					
Inlet Crude, kbpd	175					
Water (vol%)	4.8	7	10	15	20	317
Water, kbpd	8.8	13.2	19.4	30.9	43.8	317
HX-220 Duty, MW	18.70	19.76	21.24	23.99	27.07	45
V-220 High Pressure Vessel Gas Production, MMSCFD	14.94	14.99	15.05	15.15	15.24	2

Table 23 & 24 shows the effect of increasing water volume at constant outlet temperature and inlet crude flow rate for both HYSYS Simulation methods (Pseudo Components & Crude Assays). By increasing 317% of water volume in the inlet crude from 4.8 to 20 vol%, the duty requirement of HX-220X will increase by 43 - 45%. This yields a ratio of 1 : 0.142 $\pm$.

4.9.2 Scenario 2 — Variable Inlet Crude Flow at Constant Outlet Temperature & Water Volume %

Table 25 : Scenario 2 for HYSYS Simulation Pseudo Component's Method

Scenario 2 (Pseudo Comp.)	1	2	3	4	5	% Deviation
HX-220 Inlet Temperature	57					
HX-220 Outlet Temperature	87					
Inlet Crude, kbpd	150	160	170	180	190	27
Water (vol%)	4.8					
Water, kbpd	7.56	8.07	8.57	9.08	9.58	27
HX-220 Duty, MW	16.80	17.92	19.04	20.16	21.28	27
V-220 High Pressure Vessel Gas Production, MMSCFD	10.89	11.62	12.35	13.07	13.80	27

Table 26 : Scenario 2 for HYSYS Simulation Crude Assay's Method

Scenario 2 (Crude Assays)	1	2	3	4	5	% Deviation
HX-220 Inlet Temperature	57					
HX-220 Outlet Temperature	87					
Inlet Crude, kbpd	150	160	170	180	190	27
Water (vol%)	4.8					
Water, kbpd	7.56	8.07	8.57	9.08	9.58	27
HX-220 Duty, MW	16.03	17.11	18.17	19.24	20.31	27
V-220 High Pressure Vessel Gas Production, MMSCFD	12.80	13.66	14.51	15.36	16.22	27

Table 25 & 26 shows the effect of increasing crude flow at constant outlet temperature and constant water volume% for both HYSYS Simulation methods (Pseudo Components & Crude Assays). By increasing 27% of inlet crude (from 130 to 180 kbpd), both the actual water volume and duty requirement will also increase by 27%. This yield a ratio of 1 : 1.

4.9.3 Scenario 3 — Variable Water Volume % at Constant Inlet Crude Flow & Heat Exchanger Duty

Table 27 : Scenario 3 for HYSYS Simulation Pseudo Component's Method

Scenario 3 (Pseudo Comp.)	1	2	3	4	5	% Deviation
HX-220 Inlet Temperature	57					
HX-220 Outlet Temperature	90.05	88.44	86.41	83.25	80.41	-11
Inlet Crude, kbpd	175					
Water (vol%)	4.8	7	10	15	20	317
Water, kbpd	8.8	13.2	19.4	30.9	43.8	317
HX-220 Duty, MW	21.68					
V-220 High Pressure Vessel Gas Production, MMSCFD	13.53	13.12	12.63	11.91	11.31	-16

Table 28 : Scenario 3 for HYSYS Simulation Crude Assay's Method

Scenario 3 (Crude Assays)	1	2	3	4	5	% Deviation
HX-220 Inlet Temperature	57					
HX-220 Outlet Temperature	91.50	89.77	87.60	84.21	81.19	-11
Inlet Crude, kbpd	175					
Water (vol%)	4.8	7	10	15	20	317
Water, kbpd	8.8	13.2	19.4	30.9	43.8	317
HX-220 Duty, MW	21.68					
V-220 High Pressure Vessel Gas Production, MMSCFD	16.36	15.85	15.23	14.33	13.57	-17

27 & 28 shows the effect of increasing water volume % at constant inlet crude flow and heat exchanger duty for both HYSYS Simulation methods (Pseudo Components & Crude Assays). By increasing 317 % of water volume in the inlet crude from 4.8 to 20 vol %, the outlet temperature will decrease by 11 %. This yield a ratio of 1 : 0.035. Furthermore, there will be 17 % less gas produced from High Pressure Separator, V-220 A/B. As a result, the gas compressor will be running at recycle flow and increase of as flaring.

From all three scenarios, it can be concluded that **actual water volume (kbbl/d)** has greater impact on the duty of HX-220X compared to water volume %. Since the design water volume is 21 600 bbl/d (which is equivalent to 4.8 % of 450 000 bbl/d); at the average current plant production of 195 000 bbl/d. HX-220X is capable to handle water content up to 9.9 vol % (21 600 bbl/d). During pigging activities, the water content was observed to receive an average of 15 vol %, the pre-heater's capability is limited.

CHAPTER 5

CONCLUSION AND RECOMMENDATIONS

5.1 CONCLUSION

From the preliminary research, separation efficiency can be enhanced by several factors which are the separator operating pressure, and also optimum number of staged separation. In addition, fine tuning of operating condition of individual equipment should lead to high quality and maximum production of crude oil. This project aims to simulate an industrial case study which is based on Terengganu Crude Oil Terminal – TCOT operations in order to obtain a stabilized crude with maximum True Vapor Pressure of 83 kPa (12 psia) for storage/export. In addition, to propose a range of optimum operating condition that maximize the oil and gas production and study the limitation of crude oil stabilization. The inlet crude composition is taken from Tapis Blend composition at 17 barg and 30 ^{o}C.

From the research, for a stabilized crude with True Vapor Pressure of 10 - 12 psia, the incoming dry feed flow rate should be from 158 - 200 kbbl/d at pressure and temperature of 17 barg and (24 - 32 ^{o}C) respectively. Moreover, the High Pressure Separator, V-220 A/B operating pressure should be around (400 - 592 kPa) while Low Pressure Separator, V-230 A/B pressure should be set around (165 - 186 kPa). On the other hand, the outlet temperature of Hot Oil to Crude Heat Exchanger, HX-220X must be around 85 - 90 ^{o}C with three fired heater (HX-610 B/C/D) in operation.

Based on the scenario analyzed, it can be concluded that actual water volume (kbbl/d) has greater impact towards the crude oil stabilization operation. Thus, in order to obtain in specifications stabilized crude at TVP of 12 psia, incoming free water flowrate should be less than 19 kbbl/d (10 vol %) at normal production of 195 kbbl/d.

5.2 RECOMMENDATIONS

There are a few other aspects of this research that can be approached in order to improve the findings of this projects. The recommendations are as below;

a. Compare the simulation data with data from Labuan and Miri Crude (which have slightly lower quality than TAPIS Blend) to validate whether the simulation is applicable for the East Malaysia regions.
b. Another parameter that can be studied for its effects towards the product True Vapor Pressure is the sulphur content inside the incoming live crude.
c. Create a dynamic HYSYS model of fired heater (furnace system) to observe the efficiencies of each fired heater.
d. Include an economic study that would take into account the cost of utilities and processing and find the most optimum operating conditions that would results in highest gross profit.
e. Conduct another simulations with a different simulation software in order to compare its results with the current results and investigate what causes the differences.
f. Conduct a design of experiments (DOE) study that could investigate the effects of more than one parameter at once towards the product properties so as to find the optimum parameter the optimum parameters where the product of highest quality can be obtained.
g. Conduct a latest site visit to Terengganu Crude Oil Terminal, TCOT to monitor the limitation/performance of each facilities.
h. Study the differences between normal operation and during pigging activities.

REFERENCES

[1] R.C Burruss & R.T. Ryder, *Composition of Crude Oil and Natural Gas*. U.S Geological Survey, 2003.

[2] Oilfield Separation & Separators-Separator Function. (2010). Retrieved February 10, 2013, from: http://articles.compressionjobs.com/articles/oilfield

[3] GPSA Engineering Data Book, 11th ed. (Electronic)

[4] ExxonMobil Exploration and Production Malaysia Inc. (EMEPMI) General Specifications (GS)

[5] Manning & Thompson, *Oilfield Processing of Petroleum : Stabilization and Sweetening of Crude Oil,* vol. 2. PennWell Books, 1991.

[6] PETRONAS Technical Standards, Design & Engineering (Core), Rev. 12, 2004.

[7] List of Crude Oil Products, (2012). Retrieved February 10, 2013, from: http://en.wikipedia.org/wiki/List_of_crude_oil_products

[8] U.S Environmental Protection Agency, *Hazardous Air Pollutants (HAP) Specification Methodology : Emission Factor,* pg. 54.

[9] Malaysia's Environmental Law, ENVIRONMENTAL QUALITY ACT, 1974, the Malaysia Environmental Quality (Sewage and Industrial Effluents) Regulation, 1979

[10] Konstantin Panoussopoulus, *Separation of Crude Oil-Water Emulsions : Experimental Techniques and Models*, 1998.

[11] ESMAP, *Crude Oil Price Differentials and Differences in Oil Qualities : A Statistical Analysis*, 2005.

[12] KLM Technology Group, *Separator Vessel Selection and Sizing : Engineering Design Guideline*, 2011.

[13] WO 2012/127295 A1 Patent, A Process For The Recovery of Crude, 2011.

[14] Nael N. Zaki,, Ruben G. Carbonell and Peter K. Kilpatrick, *A Novel Process for Demulsification of Water-in-Crude Oil Emulsions by Dense Carbon Dioxide*, North Carolina State University, 2003.

[15] J. C. Jones, *Hydrocarbons : Physical Properties and their Relevance to Utilization,* Ventus Publishing ApS, 2010.

[16] Dr. Paul Glover, Chapter 2 : Reservoir Fluids [PowerPoint Slides]

[17] K. Arnold and M. Stewart, *Surface Production Operation : Design of Oil-Handling System and Facilities*, 2nd ed. Woburn, MA: Butterworth-Heinemann, 1999, vol 1.

[18] Manish V.Shah, *Practical Selection and Design Guide: For Gas Liquid Separators With Optimization*, 2000.

[19] Ernest W.Klechka, *Use of Corrosion Inhibitors on the Trans Alaska Pipeline*, CORTEC Corp., 2001.

[20] Hazard Devold, *Oil and Gas Production Handbook*, ED 1.3, ABB ATPA Oil and Gas, London, June 2006.

[21] Richard M. Felder, *Elementary Principles of Chemical Processes*, Third Edition, John Wiley & Sons, Inc, New York

[22] Terengganu Crude Oil Terrminal (TCOT) Start-Up and Operating Manual

[23] SKG16 Facilities Process [PowerPoint Slides]

[24] Terengganu Crude Oil Terminal Operation Training [PowerPoint Slides]

[25] Feasibility Study For Terengganu Crude Oil Terminal (TCOT) & Liquefied Petroleum Gas Unit (LPGRU) Debottlenecking Project (TCOT-PR-RPT-0001_R0)

[26] Terengganu Crude Oil Terminal (TCOT) Design Basis Memorandum

[27] Norrie. (2010), Compression Principle & Gas Compressors — Surging in a Centrifugal Compressor. Retrieved February 15, 2013, from: http://articles.compressionjobs.com

[28] MMC Oil & Gas Engineering Sdn Bhd. (2011), TCOT Incoming Fluid Composition – Separation Improvement and Sand Removal Installation Report.

[29] Francis S. Manning, Richard E. hompson, *Chapter 3: Phase Behavior, Oilfield Processing of Petroleum : Crude Oil*, Vol. 2. 1995.

[30] Gerard Chin, Whole Crude Compositional Analysis, Laboratory Analysis Report for PETRONAS Carigali Sdn Bhd, 2010.

[31] Group Technical Solutions, PMO SWAT- TCOT water Fluctuation Issue — Phase Repor (Process Engineering), 2010.

APPENDIX I

RELATED DRAWING

TYPICAL CRUDE STABILIZATION

TCOT – HYSYS MODEL

TCOT – PFD

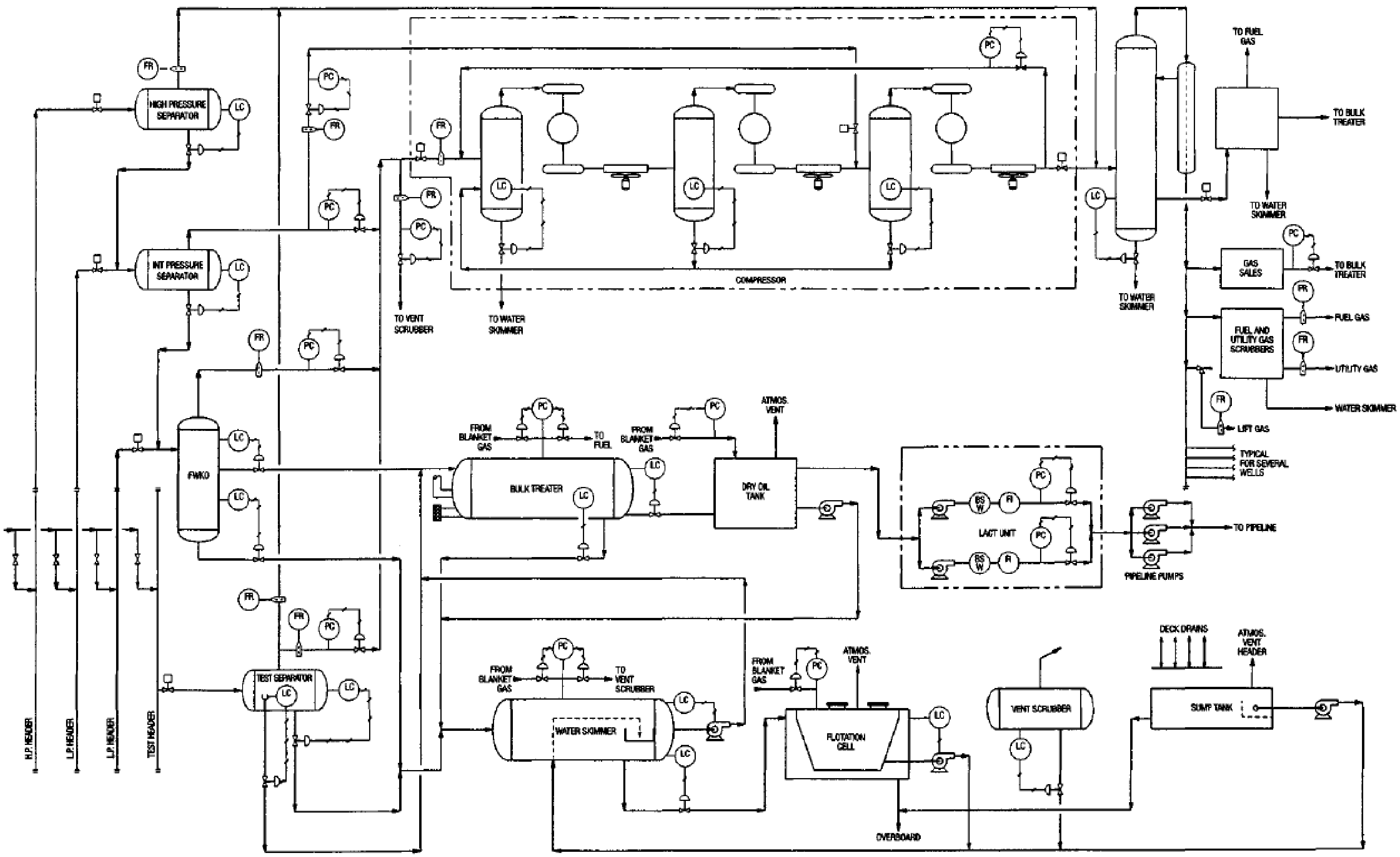

Figure 2 (Section 2.2 Oil Handling Facilities, pg. 09) : Typical PFD of Onshore Crude Oil Receiving Terminal

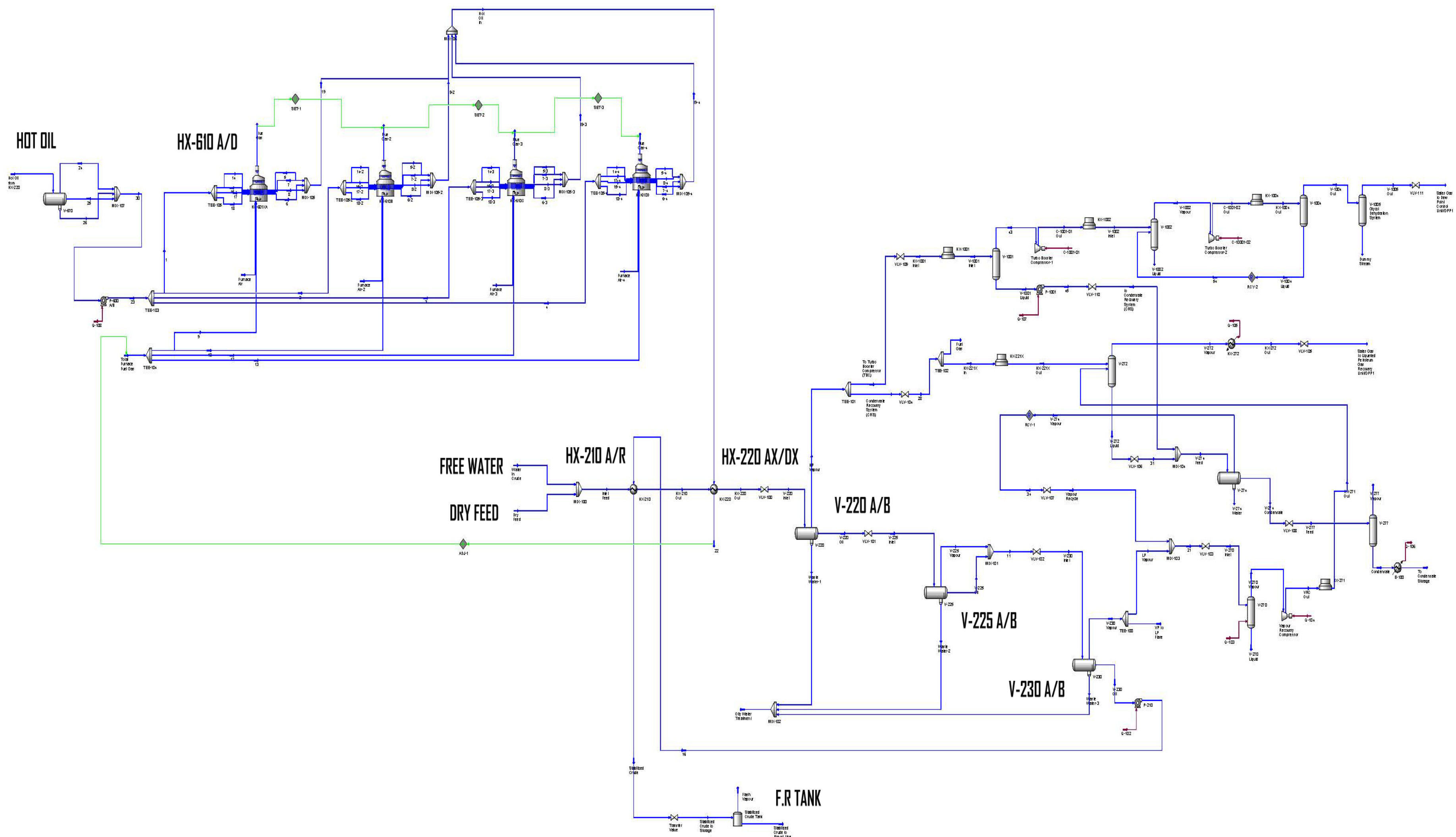

Figure 9 (Section 4.1 Brief Process Description, pg. 27) : PFD of TCOT Crude Stabilization HYSYS Model

ITEM	HX 210 A-R	HX 220 A-X	V 270 A/B	HX 1001X	V 1001	V 270 A-D	HX 1002 A/B	V 1002	HX 1004	V 1004	F 1002 A/B	V 1005	HX 271X	IV 225 A/B	V 230 A/B	HX 271 A-D	V 281	V 272	V 274	V 277
NAME	CRUDE TO CRUDE EXCHANGER	HOT OIL TO CRUDE EXCHANGER	GAS OIL WATER SEPARATOR	SUCTION COOLER	SUCTION SCRUBBER	VAPOR RECOVERY COMPRESSOR	FIRST STAGE AFTER COOLER	SECOND STAGE AFTER COOLER	DISCHARGE COOLER	DISCHARGE SCRUBBER	CONDENSATE FILTER SEPARATOR	GLYCOL CONTACTOR	HP OFF-GAS COOLER	ELECTROSTATIC PRECIPITATOR	CRUDE SURGE SEPARATOR	AFTER COOLER	CONDENSATE STABILIZER	DISCHARGE SEPARATOR	LP CONDENSATE RECOVERY SEPARATOR	ATHMOSPHERIC CONDENSATE RECOVERY SEPARATOR

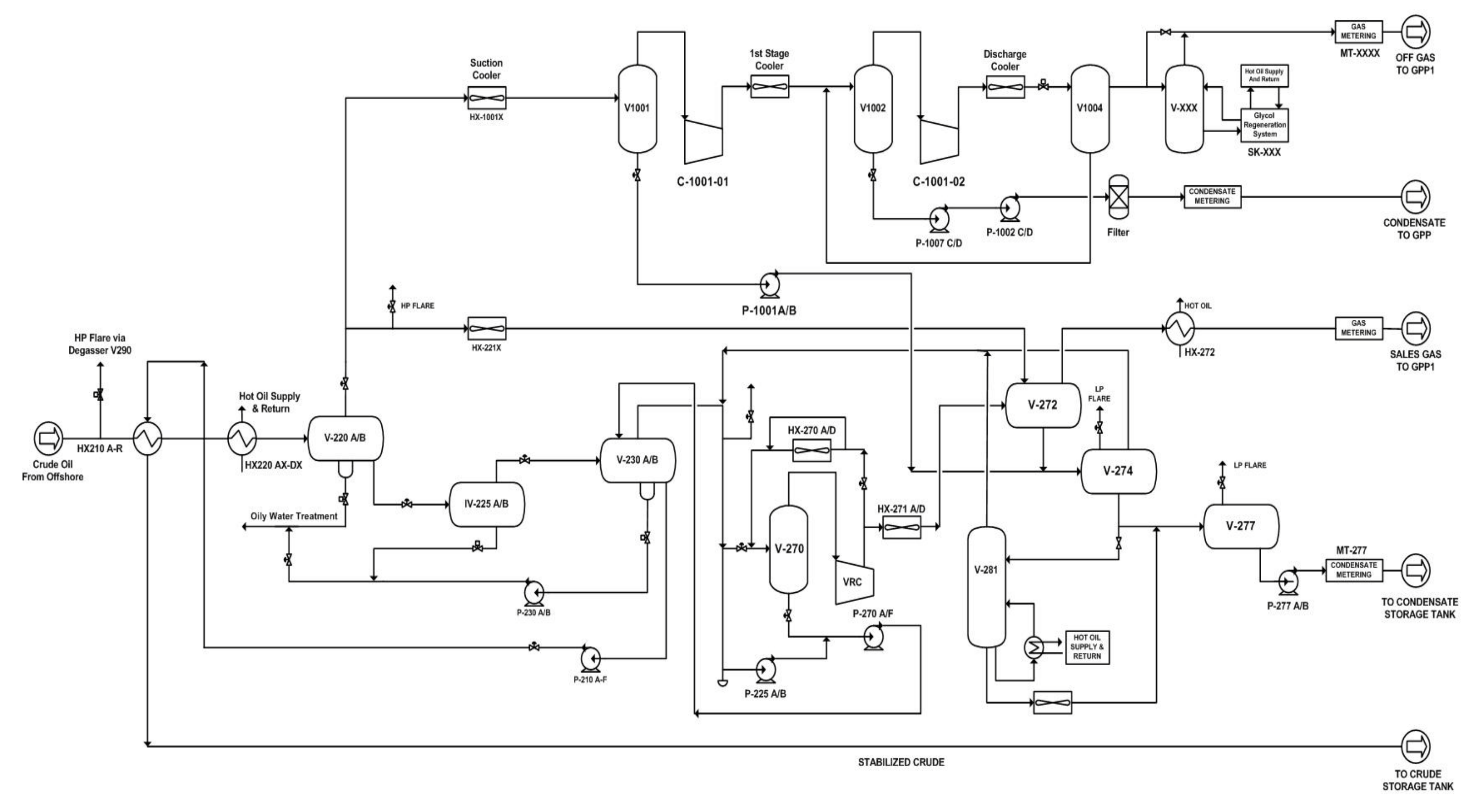

Figure 29 : PFD of Crude Oil Stabilization System at Terengganu Crude Oil Terminal, TCOT

APPENDIX II

RELATED DRAWING

TCOT CRUDE STABILIZATION

SYSTEM

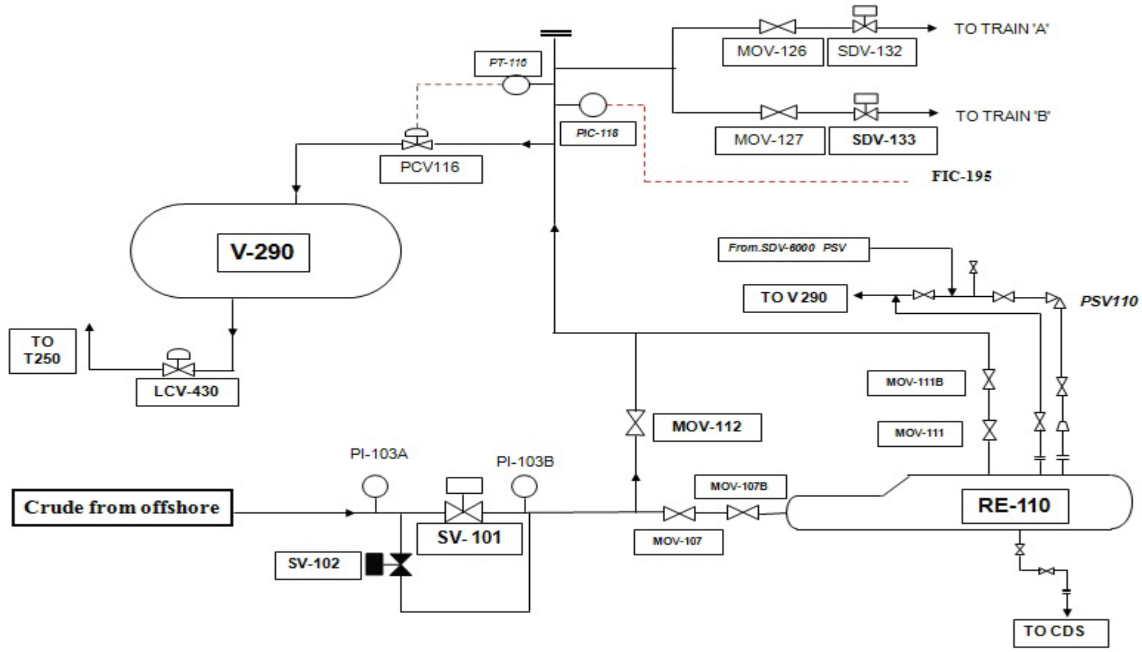

Figure 30 : TCOT Crude Receiving System

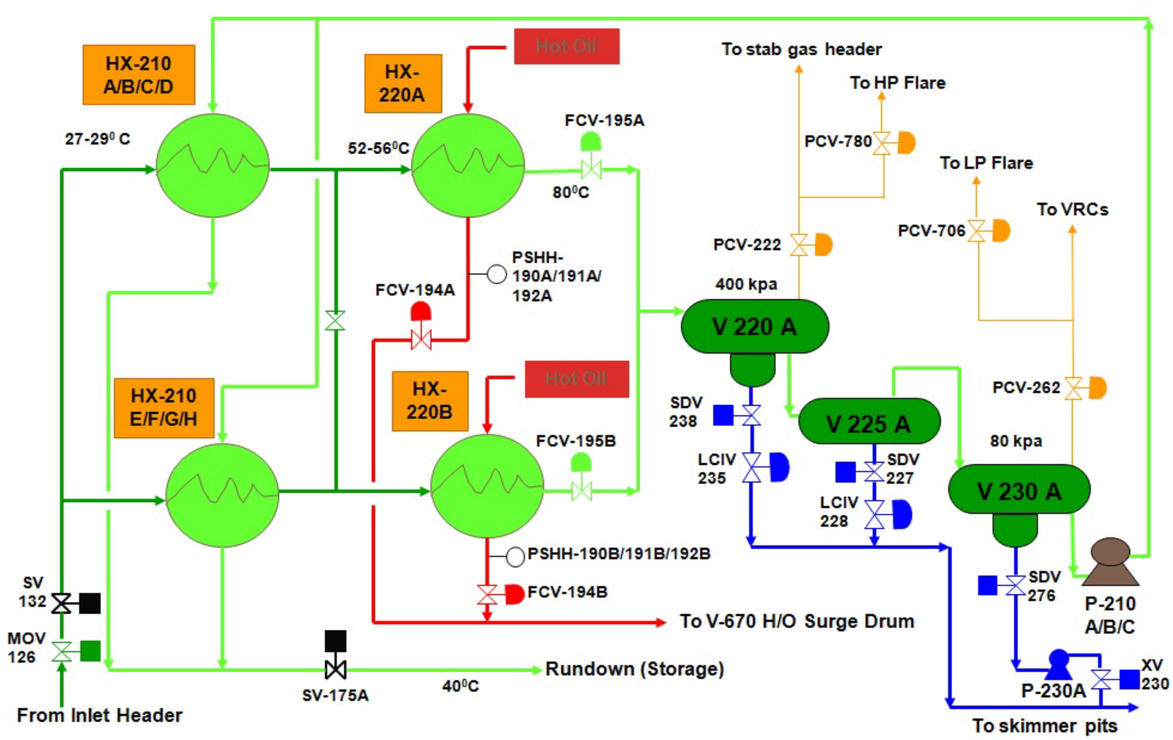

Figure 11 : TCOT Crude Stabilization System

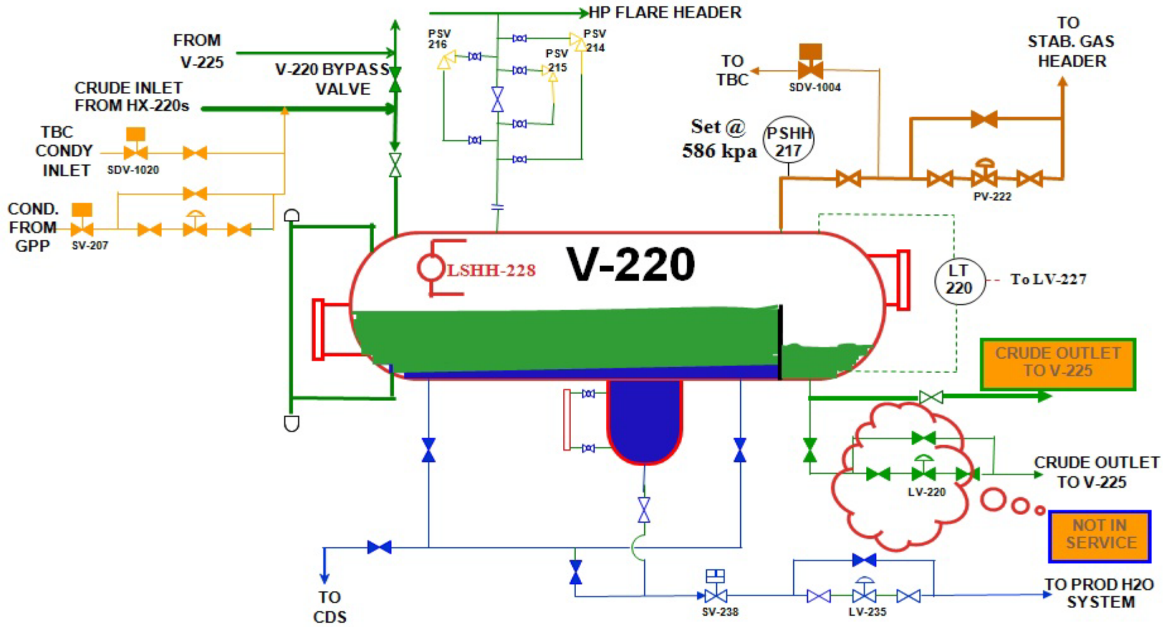

Figure 32 : Gas Oil Water Separator V-220

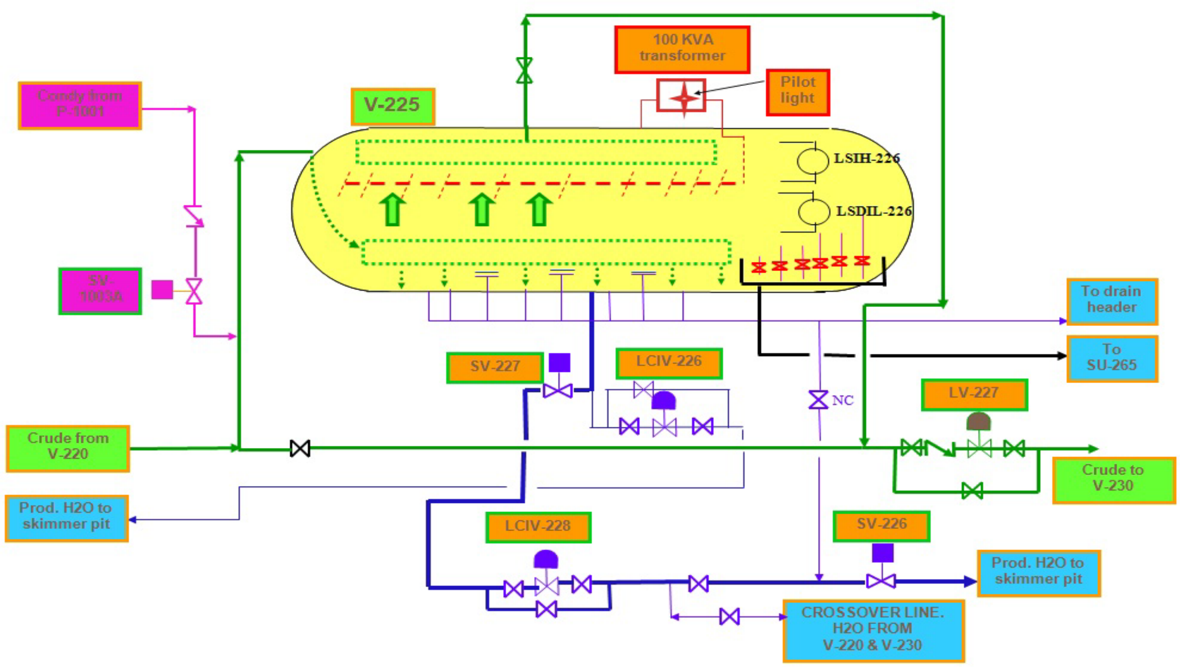

Figure 33 : Electrostatic Precipitator V-225

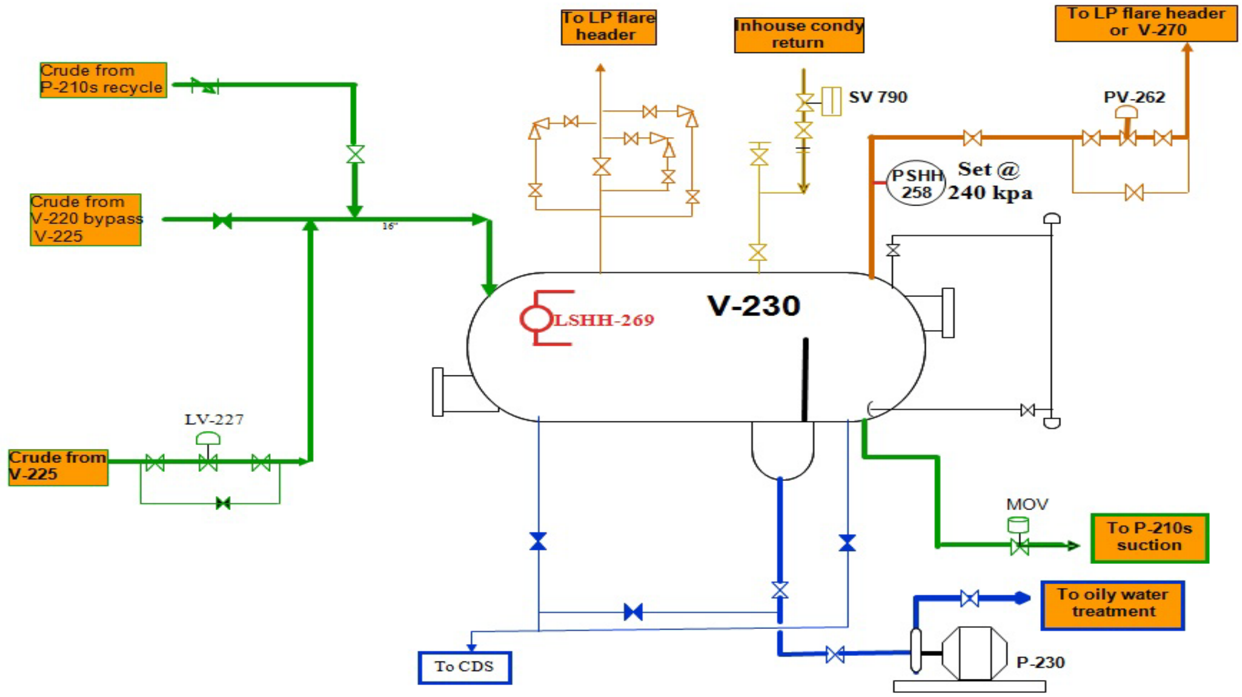

Figure 34 : Crude Surge Separator V-230

APPENDIX III

BASIS OF SIMULATIONS

INLET CRUDE COMPOSITION

TCOT PLANT DATA

INCOMING CRUDE FROM TAPIS PUMP DRY BASIS

	Mol. Weight	Formula	Components	Mol %	Mol Fraction	Mass, kg	Mass Fraction	Mass Flow, kg/hr	Molar Flowrate, kmol/hr
1	2.016	H2	Hydrogen	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
2	34.08	H2S	Hydrogen Sulphide	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
3	44.01	CO2	Carbon Dioxide	0.5067	0.0051	0.2229	0.0014	1298.0506	29.4944
4	28.01	N2	Nitrogen	0.0300	0.0003	0.0084	0.0001	48.9129	1.7463
5	16.04	C1	Methane	4.1400	0.0414	0.6637	0.0041	3865.3966	240.9848
6	30.07	C2	Ethane	2.5000	0.0250	0.7514	0.0047	4375.8537	145.5222
7	44.1	C3	Propane	4.0000	0.0400	1.7632	0.0109	10268.0491	232.8356
8	58.12	i-C4	i-Butane	2.4600	0.0246	1.4291	0.0089	8322.4284	143.1939
9	58.12	n-C4	n-Butane	2.9200	0.0292	1.6963	0.0105	9878.6549	169.9700
10	72.15	Neo-C5	Neo-Pentane	0.0100	0.0001	0.0072	0.0000	41.9977	0.5821
11	72.15	i-C5	i-Pentane	2.9700	0.0297	2.1419	0.0133	12473.3222	172.8804
12	72.15	n-C5	n-Pentane	2.1100	0.0211	1.5217	0.0094	8861.5185	122.8208
13	86.18	C6	Hexane	5.0100	0.0501	4.3156	0.0268	25132.3774	291.6266
14	84.16		M-Cyclo Pentane	1.3400	0.0134	1.1272	0.0070	6564.4732	77.9999
15	78.11		Benzene	0.2500	0.0025	0.1952	0.0012	1136.6742	14.5522
16	84.16		Cyclo-hexane	0.9900	0.0099	0.8328	0.0052	4849.8720	57.6268
17	100.2	C7	Heptane	4.6300	0.0463	4.6371	0.0288	27004.6200	269.5072
18	98.19		M-C-Hexane	2.6400	0.0264	2.5910	0.0161	15089.0030	153.6715
19	92.14		Toluene	1.1700	0.0117	1.0775	0.0067	6275.1401	68.1044
20	114.2	C8	Octanes	5.9100	0.0591	6.7461	0.0419	39286.4640	344.0146
21	106.2		E-Benzene	0.3300	0.0033	0.3503	0.0022	2039.9889	19.2089
22	106.2		M/P-Xylene	1.8100	0.0181	1.9213	0.0119	11189.0303	105.3581
23	106.2		O-Xylene	0.4800	0.0048	0.5095	0.0032	2967.2566	27.9403
24	128.3	C9	Nonanes	4.3800	0.0438	5.6169	0.0349	32710.7215	254.9550
25	120.2		1,2,4-TMB	0.6700	0.0067	0.8050	0.0050	4687.7952	39.0000
26	142.3	C10	Decanes	4.9400	0.0494	7.0263	0.0436	40918.6414	287.5519
27	156.3	C11	Undecanes	4.7100	0.0471	7.3583	0.0457	42851.8171	274.1639
28	170.3	C12	Dodecanes	4.1400	0.0414	7.0471	0.0437	41039.7159	240.9848
29	184.4	C13	Tridecanes	4.2300	0.0423	7.7965	0.0484	45403.6368	246.2236
30	198.4	C14	Tetradecanes	3.8000	0.0380	7.5357	0.0468	43884.8503	221.1938
31	212.4	C15	Pentadecanes	4.0500	0.0405	8.5982	0.0534	50072.4559	235.7460
32	226.4	C16	Hexadecanes	2.9200	0.0292	6.6078	0.0410	38481.2022	169.9700
33	240.5	C17	Heptadecanes	2.4400	0.0244	5.8655	0.0364	34158.1439	142.0297
34	254.5	C18	Octadecanes	2.4200	0.0242	6.1560	0.0382	35850.2765	140.8655
35	268.5	C19	Nonadecanes	1.9100	0.0191	5.1260	0.0318	29851.5588	111.1790
36	282.5	C20	Eicosanes	1.5600	0.0156	4.4049	0.0273	25652.6602	90.8059
37	296.6	C21	Heneicosanes	1.3700	0.0137	4.0615	0.0252	23652.7189	79.7462
38	310.6	C22	Docosanes	1.1900	0.0119	3.6944	0.0229	21514.8226	69.2686
39	324.6	C23	Tricosanes	1.0300	0.0103	3.3418	0.0207	19461.4456	59.9552
40	338.6	C24	Tetracosanes	0.9200	0.0092	3.1137	0.0193	18132.7694	53.5522
41	352.7	C25	Pentacosanes	0.8100	0.0081	2.8555	0.0177	16629.5247	47.1492
42	366.7	C26	Hexacosanes	0.7200	0.0072	2.6390	0.0164	15368.5454	41.9104
43	380.7	C27	Heptacosanes	0.6700	0.0067	2.5495	0.0158	14847.2847	39.0000
44	394.7	C28	Octacosanes	0.6000	0.0060	2.3671	0.0147	13785.0306	34.9253
45	408.8	C29	Nonacosanes	0.5800	0.0058	2.3699	0.0147	13801.5619	33.7612
46	422.8	C30	Tricontanes	0.5100	0.0051	2.1553	0.0134	12551.4677	29.6865
47	435.3	C31	Hentriacontanes	0.4400	0.0044	1.9144	0.0119	11148.8661	25.6119
48	450.9	C32	Dotriacontanes	0.3400	0.0034	1.5323	0.0095	8923.7729	19.7910
49	464.3	C33	Tritriacontanes	0.2900	0.0029	1.3458	0.0084	7837.6531	16.8806
50	478.3	C34	Tetratriacontanes	0.2200	0.0022	1.0518	0.0065	6125.0892	12.8060
51	492.4	C35	Pentatriacontanes	0.1900	0.0019	0.9351	0.0058	5445.7914	11.0597
52	599.7	C36+	Hexatriacontanes Plus	1.7900	0.0179	10.7296	0.0666	62485.0954	104.1939
				100.0467	1.0000	161.1104	1.0000	938244.0000	5823.6079

STABILIZED CRUDE (TCOT EXPORT LINE) DRY BASIS

	Formula	Components	Mol %	Mol Fraction	Wt %
1	N2	N2	0.0190	0.0002	0.0030
2	CO2	CO2	0.0000	0.0000	0.0000
3	C1	C1	0.0020	0.0000	0.0000
4	C2	C2	0.0260	0.0003	0.0050
5	C3	C3	0.4940	0.0049	0.1240
6	i-C4	i-C4	1.4840	0.0148	0.4930
7	C4	C4	0.8410	0.0084	0.2790
8		Neo-Pentane	0.0000	0.0000	0.0000
9	i-C5	i-C5	3.0550	0.0305	1.2600
10	C5	C5	2.2990	0.0230	0.9480
11		2,2-Dimethylbutane*	0.0000	0.0000	0.0000
12		2,3-Dimethylbutane*	0.0000	0.0000	0.0000
13		Cyclopentane	0.2760	0.0028	0.1340
14		2-Methylpentane	2.2080	0.0221	1.0750
15		3-Methylpentane	1.1600	0.0116	0.5650
16	C6	C6	2.2890	0.0229	1.1140
17		Methylcyclopentane	1.1200	0.0112	0.5890
18		2,4-dimethylpentane	0.3550	0.0035	0.1860
19		Benzene	0.2700	0.0027	0.1420
20		Cyclohexane	1.0260	0.0103	0.5400
21		2-Methylhexane*	0.0000	0.0000	0.0000
22		3-Methylhexane*	0.0000	0.0000	0.0000
23		1,T-3-Dimethylcyclopentane*	0.0000	0.0000	0.0000
24		1,C-3-Dimethylcyclopentane*	0.0000	0.0000	0.0000
25		1,T-2-Dimethylyclopentane*	0.0000	0.0000	0.0000
26	C7	C7	6.4120	0.0641	3.3720
27		Methylcyclohexane	2.8590	0.0286	1.7030
28		Ethylcyclopentane*	0.0000	0.0000	0.0000
29		Toluene	1.2570	0.0126	0.7480
30		2-Methylheptane*	0.0000	0.0000	0.0000
31		1-Cis-3-Dimethylcyclohexane*	0.0000	0.0000	0.0000
32		3-Methylheptane*	0.0000	0.0000	0.0000
33		1-Methyl-T-3-Ethylcyclopentane*	0.0000	0.0000	0.0000
34		1-Trans-2-Dimethylcyclohexane*	0.0000	0.0000	0.0000
35	C8	C8	8.4520	0.0845	5.0340
36		Ethylcyclohexane*	0.0000	0.0000	0.0000
37		Ethylbenzene	0.1270	0.0013	0.0880
38		Meta & Para-Xylene	1.9230	0.0192	1.3260
39		Ortho-Xylene	0.6760	0.0068	0.4660
40		2-Methyloctane*	0.0000	0.0000	0.0000
41		3-Methyloctane*	0.0000	0.0000	0.0000
42	C9	C9	5.7300	0.0573	3.9530
43		1-Methyl-3-Ethylbenzene*	0.0000	0.0000	0.0000
44		1,3,5-Trimethylbenzene*	0.0000	0.0000	0.0000
45		n-Propylbenzene*	0.0000	0.0000	0.0000
46		1-Methyl-2-Ethylbenzene*	0.0000	0.0000	0.0000
47	C10	C10	7.0980	0.0710	5.4350
48		n-Butylbenzene*	0.3960	0.0040	0.3320
49	C11	C11	5.0470	0.0505	4.2390
50	C12	C12	4.7540	0.0475	4.3740
51	C13	C13	4.7300	0.0473	4.7300
52	C14	C14	3.9840	0.0398	4.3250
53	C15	C15	4.3810	0.0438	5.1570
54	C16	C16	3.0570	0.0306	3.8790
55	C17	C17	2.5560	0.0256	3.4620
56	C18	C18	2.5330	0.0253	3.6340
57	C19	C19	1.9940	0.0199	2.9960
58	C20	C20	1.6130	0.0161	2.5340
59	C21	C21	1.5350	0.0153	2.5530
60	C22	C22	1.2680	0.0127	2.2090
61	C23	C23	1.1510	0.0115	2.0920
62	C24	C24	1.0310	0.0103	1.9490
63	C25	C25	0.9890	0.0099	1.9500
64	C26	C26	0.9260	0.0093	1.9010
65	C27	C27	0.6700	0.0067	1.4330
66	C28	C28	0.7590	0.0076	1.6830
67	C29	C29	0.6950	0.0069	1.5970
68	C30	C30	0.5450	0.0054	1.2950
69	C31	C31	0.4590	0.0046	1.1280
70	C32	C32	0.4110	0.0041	1.0430
71	C33	C33	0.3290	0.0033	0.8600
72	C34	C34	0.2790	0.0028	0.7530
73	C35	C35	0.2100	0.0021	0.5820
74	C36	C36	0.2040	0.0020	0.6130
75	C37	C37	0.1430	0.0014	0.4190
76	C38	C38	0.1350	0.0013	0.4060
77	C39	C39	0.1050	0.0010	0.3240
78	C40+	C40+	1.6550	0.0165	5.9660
	TOTAL		100.0020	1.0000	100.0000

APPENDIX IV

PRELIMINARY ANALYSIS

TABLE DATA

TCOT PLANT DATA vs. HYSYS PSEUDO COMPONENT (Figure 12, pg 37)

	DRY BASIS COMPOSITION		TCOT Real Plant Data	HYSYS Pseudo Component
	Formula	**Components**	Mol Fraction	Mol Fraction
1	N2	Nitrogen	0.000190	0.000001
2	C1	Methane	0.000020	0.000511
3	C2	Ethane	0.000260	0.003122
4	C3	Propane	0.004940	0.015277
5	i-C4	i-Butane	0.014840	0.015689
6	n-C4	n-Butane	0.008410	0.020863
7	Neo-C5	Neo-Pentane	0.000000	0.000080
8	i-C5	i-Pentane	0.030549	0.027002
9	n-C5	n-Pentane	0.022990	0.019985
10	C6	Hexane	0.022890	0.053485
11	C7	Heptane	0.064119	0.051893
12	C8	Octanes	0.084518	0.067591
13	C9	Nonanes	0.057299	0.050505
14	C10	Decanes	0.070979	0.057159
15	C11	Undecanes	0.050469	0.054578
16	C12	Dodecanes	0.047539	0.048001
17	C13	Tridecanes	0.047299	0.049059
18	C14	Tetradecanes	0.039839	0.044078
19	C15	Pentadecanes	0.043809	0.046979
20	C16	Hexadecanes	0.030569	0.033872
21	C17	Heptadecanes	0.025559	0.028304
22	C18	Octadecanes	0.025329	0.028072
23	C19	Nonadecanes	0.019940	0.022156
24	C20	Eicosanes	0.016130	0.018096
25	C21	Heneicosanes	0.015350	0.015892
26	C22	Docosanes	0.012680	0.013804
27	C23	Tricosanes	0.011510	0.011948
28	C24	Tetracosanes	0.010310	0.010672
29	C25	Pentacosanes	0.009890	0.009396
30	C26	Hexacosanes	0.009260	0.008352
31	C27	Heptacosanes	0.006700	0.007772
32	C28	Octacosanes	0.007590	0.006960
33	C29	Nonacosanes	0.006950	0.006728
34	C30	Tricontanes	0.005450	0.005916
35	C31*	Hentriacontanes	0.004590	0.005104
36	C32	n-Dotriacontanes	0.004110	0.003944
37	C33*	Tritriacontanes	0.003290	0.003364
38	C34*	Tetratriacontanes	0.002790	0.002552
39	C35*	Pentatriacontanes	0.002100	0.002204
40	C36*+	Hexatriacontanes Plus	0.022420	0.020764
41	M-Cyclo Pentane	M-Cyclo Pentane	0.051189	0.014288
42	Benzene	Benzene	0.002700	0.002653
43	Cyclo-hexane	Cyclo-hexane	0.010260	0.010689
44	M-C-Hexane	M-C-Hexane	0.028589	0.029490
45	Toluene	Toluene	0.012570	0.013112
46	E-Benzene	E-Benzene	0.001270	0.003773
47	M/P-Xylene	M/P-Xylene	0.019230	0.020729
48	O-Xylene	O-Xylene	0.006760	0.005501
49	1,2,4-TMB	1,2,4-TMB	0.003960	0.007737
50	CO2	Carbon Dioxide	0.000000	0.000295
51	H20	Water	0.000000	0.000000
			1.000000	1.000000

TOTAL VALIDATION, PLANT DATA vs. HYSYS SIMULATIONS (Fig 14, pg 40)

	DRY BASIS COMPOSITION		TCOT Real Plant Data	HYSYS Pseudo Component	HYSYS Crude Assay
	Formula	**Components**	**Mol Fraction**	**Mol Fraction**	**Mol Fraction**
1	N2	Nitrogen	0.000190	0.000001	0.000000
2	C1	Methane	0.000020	0.000511	0.000373
3	C2	Ethane	0.000260	0.003122	0.002895
4	C3	Propane	0.004940	0.015277	0.015199
5	i-C4	i-Butane	0.014840	0.015689	0.015622
6	n-C4	n-Butane	0.008410	0.020863	0.021200
7	Neo-C5	Neo-Pentane	0.000000	0.000080	0.000079
8	i-C5	i-Pentane	0.030549	0.027002	0.027257
9	n-C5	n-Pentane	0.022990	0.019985	0.020263
10	C6*+	Hexane & Heavier	0.917802	0.897176	0.896866
11	CO2	Carbon Dioxide	0.000000	0.000295	0.000245
		TOTAL	1.000000	1.000000	1.000000

HYSYS PSEUDO COMPONENT, INLET vs. OUTLET (Figure 15, pg 43)

		Inlet Flow	Stabilized Crude
Formula	**Components**	**Mol Fraction**	**Mol Fraction**
H2	Hydrogen	0.000000	0.000000
H2S	Hydrogen Sulphide	0.000000	0.000000
CO2	Carbon Dioxide	0.002333	0.000295
N2	Nitrogen	0.000137	0.000001
C1	Methane	0.018942	0.000510
C2	Ethane	0.011438	0.003119
C3	Propane	0.018301	0.015265
i-C4	i-Butane	0.011255	0.015676
n-C4	n-Butane	0.013360	0.020846
Neo-C5	Neo-Pentane	0.000046	0.000080
i-C5	i-Pentane	0.013589	0.026981
n-C5	n-Pentane	0.009654	0.019969
C6	Hexane	0.022922	0.053442
	M-Cyclo Pentane	0.006131	0.014277
	Benzene	0.001144	0.002651
	Cyclo-hexane	0.004530	0.010680
C7	Heptane	0.021183	0.051852
	M-C-Hexane	0.012079	0.029467
	Toluene	0.005353	0.013102
C8	Octanes	0.027040	0.067538
	E-Benzene	0.001510	0.003770
	M/P-Xylene	0.008281	0.020712
	O-Xylene	0.002196	0.005496
C9	Nonanes	0.020040	0.050465
	1,2,4-TMB	0.003065	0.007731
C10	Decanes	0.022602	0.057113
C11	Undecanes	0.021549	0.054535
C12	Dodecanes	0.018942	0.047963
C13	Tridecanes	0.019353	0.049020
C14	Tetradecanes	0.017386	0.044043
C15	Pentadecanes	0.018530	0.046942
C16	Hexadecanes	0.013360	0.033845
C17	Heptadecanes	0.011164	0.028282
C18	Octadecanes	0.011072	0.028050
C19	Nonadecanes	0.008739	0.022139
C20	Eicosanes	0.007137	0.018082
C21	Heneicosanes	0.006268	0.015880
C22	Docosanes	0.005445	0.013793
C23	Tricosanes	0.004713	0.011939
C24	Tetracosanes	0.004209	0.010664
C25	Pentacosanes	0.003706	0.009389
C26	Hexacosanes	0.003294	0.008346
C27	Heptacosanes	0.003065	0.007766
C28	Octacosanes	0.002745	0.006955
C29	Nonacosanes	0.002654	0.006723
C30	Tricontanes	0.002333	0.005911
C31*	Hentriacontanes	0.002013	0.005100
C32	n-Dotriacontanes	0.001556	0.003941
C33*	Tritriacontanes	0.001327	0.003361
C34*	Tetratriacontanes	0.001007	0.002550
C35*	Pentatriacontanes	0.000869	0.002202
C36*+	Hexatriacontanes Plus	0.008190	0.020748
H20	Water	0.542245	0.000796
		1.000000	1.000000

HYSYS PSEUDO COMPONENT, INLET vs. OUTLET DRY BASIS (Fig 16, pg 44)

Formula	Components	Inlet Crude	Stabilized Crude
H2	Hydrogen	0.000000	0.0000
H2S	Hydrogen Sulphide	0.000000	0.0000
CO2	Carbon Dioxide	0.005097	0.0003
N2	Nitrogen	0.000300	0.0000
C1	Methane	0.041379	0.0005
C2	Ethane	0.024988	0.0031
C3	Propane	0.039980	0.0153
i-C4	i-Butane	0.024588	0.0157
n-C4	n-Butane	0.029185	0.0209
Neo-C5	Neo-Pentane	0.000100	0.0001
i-C5	i-Pentane	0.029685	0.0270
n-C5	n-Pentane	0.021089	0.0200
C6	Hexane	0.050075	0.0535
	M-Cyclo Pentane	0.013393	0.0143
	Benzene	0.002499	0.0027
	Cyclo-hexane	0.009895	0.0107
C7	Heptane	0.046277	0.0519
	M-C-Hexane	0.026387	0.0295
	Toluene	0.011694	0.0131
C8	Octanes	0.059070	0.0676
	E-Benzene	0.003298	0.0038
	M/P-Xylene	0.018091	0.0207
	O-Xylene	0.004798	0.0055
C9	Nonanes	0.043778	0.0505
	1,2,4-TMB	0.006697	0.0077
C10	Decanes	0.049375	0.0572
C11	Undecanes	0.047076	0.0546
C12	Dodecanes	0.041379	0.0480
C13	Tridecanes	0.042279	0.0491
C14	Tetradecanes	0.037981	0.0441
C15	Pentadecanes	0.040480	0.0470
C16	Hexadecanes	0.029185	0.0339
C17	Heptadecanes	0.024388	0.0283
C18	Octadecanes	0.024188	0.0281
C19	Nonadecanes	0.019090	0.0222
C20	Eicosanes	0.015592	0.0181
C21	Heneicosanes	0.013693	0.0159
C22	Docosanes	0.011894	0.0138
C23	Tricosanes	0.010295	0.0119
C24	Tetracosanes	0.009195	0.0107
C25	Pentacosanes	0.008096	0.0094
C26	Hexacosanes	0.007196	0.0084
C27	Heptacosanes	0.006697	0.0078
C28	Octacosanes	0.005997	0.0070
C29	Nonacosanes	0.005797	0.0067
C30	Tricontanes	0.005097	0.0059
C31*	Hentriacontanes	0.004398	0.0051
C32	n-Dotriacontanes	0.003398	0.0039
C33*	Tritriacontanes	0.002899	0.0034
C34*	Tetratriacontanes	0.002199	0.0026
C35*	Pentatriacontanes	0.001899	0.0022
C36*+	Hexatriacontanes Plus	0.017891	0.0208
		1.000000	1.0000

HYSYS CRUDE ASSAY, INLET vs. OUTLET (Figure 17, pg 46)

		Inlet Flow	Stabilized Crude
No.	Components	Mol Fraction	Mol Fraction
1	Hydrogen	0.000000	0.000000
2	Hydrogen Sulphide	0.000000	0.000000
3	Carbon Dioxide	0.002450	0.000245
4	Nitrogen	0.000145	0.000000
5	Methane	0.020018	0.000373
6	Ethane	0.012088	0.002893
7	Propane	0.019341	0.015188
8	i-Butane	0.011895	0.015610
9	n-Butane	0.014119	0.021184
10	Neo-Pentane	0.000048	0.000079
11	i-Pentane	0.014361	0.027237
12	n-Pentane	0.010202	0.020248
13	NBP[0]43*	0.011555	0.023890
14	NBP[0]57*	0.010617	0.022982
15	NBP[0]70*	0.012442	0.027816
16	NBP[0]85*	0.021158	0.048518
17	NBP[0]100*	0.025856	0.060326
18	NBP[0]117*	0.023148	0.054715
19	NBP[0]129*	0.015839	0.037648
20	NBP[0]146*	0.020609	0.049242
21	NBP[0]158*	0.012217	0.029255
22	NBP[0]173*	0.029842	0.071590
23	NBP[0]188*	0.014135	0.033947
24	NBP[0]203*	0.015755	0.037859
25	NBP[0]217*	0.015594	0.037485
26	NBP[0]232*	0.015599	0.037504
27	NBP[0]246*	0.016323	0.039249
28	NBP[0]260*	0.016082	0.038672
29	NBP[0]275*	0.013129	0.031572
30	NBP[0]289*	0.011920	0.028665
31	NBP[0]304*	0.012228	0.029406
32	NBP[0]318*	0.011605	0.027908
33	NBP[0]332*	0.008665	0.020838
34	NBP[0]347*	0.007694	0.018502
35	NBP[0]362*	0.007327	0.017620
36	NBP[0]376*	0.006767	0.016274
37	NBP[0]390*	0.005854	0.014077
38	NBP[0]405*	0.005138	0.012355
39	NBP[0]420*	0.004750	0.011424
40	NBP[0]438*	0.007011	0.016861
41	H20	0.516477	0.000742
		1.000000	1.000000

HYSYS CRUDE ASSAY, INLET vs. OUTLET DRT BASIS (Figure 18, pg 47)

Components	Inlet Flow	Stabilized Crude
Hydrogen	0.000000	0.000000
Hydrogen Sulphide	0.000000	0.000000
Carbon Dioxide	0.005067	0.000245
Nitrogen	0.000300	0.000000
Methane	0.041400	0.000373
Ethane	0.025000	0.002895
Propane	0.040000	0.015199
i-Butane	0.024600	0.015622
n-Butane	0.029200	0.021200
Neo-Pentane	0.000100	0.000079
i-Pentane	0.029700	0.027257
n-Pentane	0.021100	0.020263
NBP[0]43*	0.023898	0.023907
NBP[0]57*	0.021957	0.022999
NBP[0]70*	0.025732	0.027836
NBP[0]85*	0.043759	0.048554
NBP[0]100*	0.053475	0.060371
NBP[0]117*	0.047873	0.054756
NBP[0]129*	0.032758	0.037676
NBP[0]146*	0.042622	0.049278
NBP[0]158*	0.025266	0.029277
NBP[0]173*	0.061718	0.071643
NBP[0]188*	0.029234	0.033972
NBP[0]203*	0.032584	0.037888
NBP[0]217*	0.032250	0.037513
NBP[0]232*	0.032260	0.037532
NBP[0]246*	0.033758	0.039279
NBP[0]260*	0.033259	0.038701
NBP[0]275*	0.027152	0.031596
NBP[0]289*	0.024652	0.028686
NBP[0]304*	0.025289	0.029428
NBP[0]318*	0.024001	0.027929
NBP[0]332*	0.017921	0.020854
NBP[0]347*	0.015912	0.018516
NBP[0]362*	0.015153	0.017634
NBP[0]376*	0.013995	0.016286
NBP[0]390*	0.012106	0.014088
NBP[0]405*	0.010625	0.012364
NBP[0]420*	0.009825	0.011433
NBP[0]438*	0.014500	0.016873
	1.000000	1.000000

EFFECT OF DRY FEED FLOW RATE (Figure 19, pg. 51)

Dry Feed Flowrate, kbpd VS. Product TVP, psia			
Feed Flow Percentage, %	Feed Flowrate, Kbd	HYSYS Pseudo Comp.	HYSYS Crude Assay
60	105	7.43	7.18
70	123	8.72	8.40
80	140	9.58	9.49
90	158	11.04	10.59
100	175	12.02	11.56
110	193	12.72	12.53
120	210	12.96	13.27
130	228	13.15	13.48
140	245	13.27	13.59
150	263	13.41	13.70
160	280	13.56	13.81
170	298	13.70	13.94
180	315	13.82	14.08
190	333	13.95	14.20
200	350	14.09	14.33
210	368	14.20	14.40
220	385	14.33	14.55
230	403	14.44	14.62

EFFECT OF INLET FEED TEMPERATURE (Figure 20, pg 52)

Feed Temperature, ^{0}C VS. Product TVP, psia		
Feed Temperature, ^{o}C	HYSYS Pseudo Comp.	HYSYS Crude Assay
0	18.06	17.54
4	17.07	16.54
8	16.15	15.59
12	15.16	14.63
16	14.31	13.78
20	13.42	12.93
24	12.63	12.12
28	11.8	11.37
32	11.11	10.64
36	10.42	10.03
40	9.746	9.41
44	9.177	8.832
48	8.613	8.319

EFFECTS OF FEED PRESSURE (Figure 21, pg. 53)

Feed Pressure, psia VS. Product TVP, psia		
Feed Pressure, bar	HYSYS Pseudo Comp.	HYSYS Crude Assay
6	12.050	11.52
10	12.090	11.58
14	12.060	11.59
18	12.020	11.56
22	11.980	11.53
26	11.920	11.50
30	11.880	11.47
34	11.820	11.45
38	11.780	11.39
42	11.740	11.36
46	11.740	11.33
50	11.710	11.30

EFFECT OF WATER FLOW RATE (Figure 22, pg. 54)

Water Flow Rate, kbpd VS. Product TVP, psia				
Water Flow Percentage, %	Free Water Flowrate, Kbd	Water Content, BS&W % vol.	HYSYS Pseudo Comp.	HYSYS Crude Assay
10	2	0.99%	9.84	9.38
20	4	1.96%	10.11	9.63
30	5	2.91%	10.28	9.77
40	7	3.85%	10.58	10.09
50	9	4.76%	10.85	10.39
60	11	5.66%	11.14	10.62
70	12	6.54%	11.28	10.80
80	14	7.41%	11.56	11.09
90	16	8.26%	11.78	11.35
100	18	9.09%	12.09	11.62
110	19	9.91%	12.21	11.75
120	21	10.71%	12.49	12.02
130	23	11.50%	12.75	12.27
140	25	12.28%	13.00	12.57
150	26	13.04%	13.11	12.69
160	28	13.79%	13.35	12.93

EFFECT OF HP SEPARATOR OPERATING PRESSURE (Figure 23, pg. 56)

V-220 Operating Pressure VS. Product TVP, psia			
Pressure Percentage, %	V220 Operating Pressure, psia	HYSYS Pseudo Comp.	HYSYS Crude Assay
70	47.31	11.74	11.78
75	50.69	11.75	11.81
80	54.07	11.79	11.82
85	57.45	11.88	11.88
90	60.83	11.91	11.91
95	64.21	11.97	11.96
100	67.59	12.02	12.00
105	70.97	12.07	12.04
110	74.35	12.12	12.08
115	77.73	12.17	12.12
120	81.11	12.22	12.16
125	84.49	12.27	12.19
130	87.87	12.32	12.22
135	91.25	12.37	12.24
140	94.63	12.45	12.26
145	98.01	12.49	12.28
150	101.39	12.53	12.30

EFFECT OF ELECTROSTATIC PRECIPITATOR PRESSURE (Figure 24, pg. 57)

V-225 Operating Pressure VS. Product TVP, psia			
Pressure Percentage, %	V220 Operating Pressure, psia	HYSYS Pseudo Comp.	HYSYS Crude Assay
70	30.46	12.02	11.56
75	32.63	12.02	11.56
80	34.81	12.02	11.56
85	36.98	12.02	11.56
90	39.16	12.02	11.56
95	41.33	12.02	11.56
100	43.51	12.02	11.56
105	45.69	12.02	11.56
110	47.86	12.02	11.56
115	50.04	12.02	11.56
120	52.21	12.02	11.56
125	54.39	12.02	11.56
130	56.56	12.02	11.56
135	58.74	12.02	11.56
140	60.91	12.02	11.56
145	63.09	12.02	11.56
150	65.27	12.02	11.56

EFFECT OF LP SEPARATOR OPERATING PRESSURE (Figure 25, pg. 58)

V-230 Operating Pressure VS. Product TVP, psia			
Pressure Percentage, %	V220 Operating Pressure, psia	HYSYS Pseudo Comp.	HYSYS Crude Assay
50	13.01	5.618	5.447
55	14.31	6.243	6.045
60	15.61	6.867	6.611
65	16.91	7.521	7.235
70	18.21	8.17	7.83
75	19.51	8.85	8.49
80	20.81	9.55	9.14
85	22.11	10.15	9.76
90	23.41	10.78	10.41
95	24.71	11.40	10.99
100	26.01	12.02	11.56
105	27.31	12.45	12.14
110	28.61	12.35	12.76
115	29.91	12.29	12.83
120	31.21	12.25	12.76
125	32.51	12.21	12.71
130	33.81	12.19	12.66
135	35.11	12.18	12.62
140	36.41	12.18	12.59
145	37.71	12.17	12.57
150	39.02	12.16	12.55

EFFECT OF HX-220s OUTLET TEMPERATURE (Figure 28, pg. 61)

HX-220 Outlet Temp. VS. Product TVP, psia		
HX-220 Outlet Temperature, ^{0}C	HYSYS Pseudo Comp.	HYSYS Crude Assay
57	19.92	19.71
61	18.76	18.41
65	17.59	17.23
69	16.53	16.02
73	15.43	15.04
77	14.39	13.94
81	13.39	12.94
85	12.48	12.00
89	11.59	11.14
93	10.76	10.33
97	9.99	10.33
101	9.28	8.86

EFFECT OF HX-210s OUTLET TEMPERATURE, (Fig 26 & 27, pg. 59)

HX-210 Outlet Temp. VS. Product TVP, psia				
HX-210 Outlet Temperature, ^0C	HX-220 Outlet Temperature, ^0C	Product TVP, psia	HX-220 Outlet Temperature, ^0C	Product TVP, psia
30	61.76	10.26	62.11	10.14
32	63.64	10.41	63.95	10.35
34	66.52	10.56	65.81	10.50
36	67.39	10.67	67.67	10.65
38	69.27	10.87	69.52	10.87
40	71.14	11.06	71.36	11.09
42	73.02	11.23	73.21	11.30
44	74.89	11.44	75.05	11.54
46	76.76	11.62	76.90	11.74
48	78.63	11.80	78.74	12.04
50	80.49	11.99	80.58	12.27
52	82.36	12.15	82.42	12.45
54	84.22	12.25	84.25	12.16
58	87.93	11.78	87.91	11.36
60	89.79	11.45	89.73	10.99
64	93.46	10.64	93.37	10.24
68	97.16	9.97	96.98	9.57
	PSEUDO COMP.		CRUDE ASSAY	

APPENDIX V

FIRED HEATER, HX-610

PERFORMANCE

CALCULATION

FIRED HEATER HX-610s PERFORMANCE CALCULATION

Fired Heater Efficiency

$$\boldsymbol{Heater\ Efficiency} = \frac{Heat\ Absorbed\ by\ Hot\ Oil\ (MW)}{Heat\ Released\ by\ Fuel\ Gas\ (MW)} \times 100\%$$

Where:

Heat Absorbed, $Q_{abs} = mC_p\Delta T$

m = Hot Oil Flow, kg/s

C_P = Average Heat Capacity of T_{in} and T_{out}, Kj/kg.K

LHV = lower heating value, KW.h/m^3

ΔT = Temperature Different

Heat Released, Q_{rel} = v $\times LHV$

v = Fuel Gas Flow, m^3/h

$$Q_{abs}\ HX610B = 186.56\ \frac{m^3}{h} \times 0.86 \times 10^3 \frac{kg}{m^3} \times 2.623\ \frac{kJ}{kg.K} \times (197.09 - 138.33)°C$$

$\boldsymbol{Q_{abs}\ HX610B}$ **= 6.91 MW**

$$Q_{abs}\ HX610C = 176.12\ \frac{m^3}{h} \times 0.86 \times 10^3 \frac{kg}{m^3} \times 2.623\ \frac{kJ}{kg.K} \times (195.25 - 138.33)°C$$

$\boldsymbol{Q_{abs}\ HX610C}$ **= 6.23 MW**

$$Q_{abs}\ HX610D = 197.0\ \frac{m^3}{h} \times 0.86 \times 10^3 \frac{kg}{m^3} \times 2.623\ \frac{kJ}{kg.K} \times (198.93 - 138.33)°C$$

$\boldsymbol{Q_{abs}\ HX610D}$ **= 7.52 MW**

Total Heat Absorbed by Hot Oil, MW = 6.91 + 6.23 + 7.52 MW = 20.75 MW

$$Q_{rel} = 123.61 \times 10^3\ \frac{m^3}{d} \times \frac{1d}{24\ h} \times 10.46\ \frac{kW.h}{m^3} = 53.87\ MW$$

Heater Efficiency = $\frac{20.75\ MW}{53.87\ MW} \times 100\% = 38.51\%$